

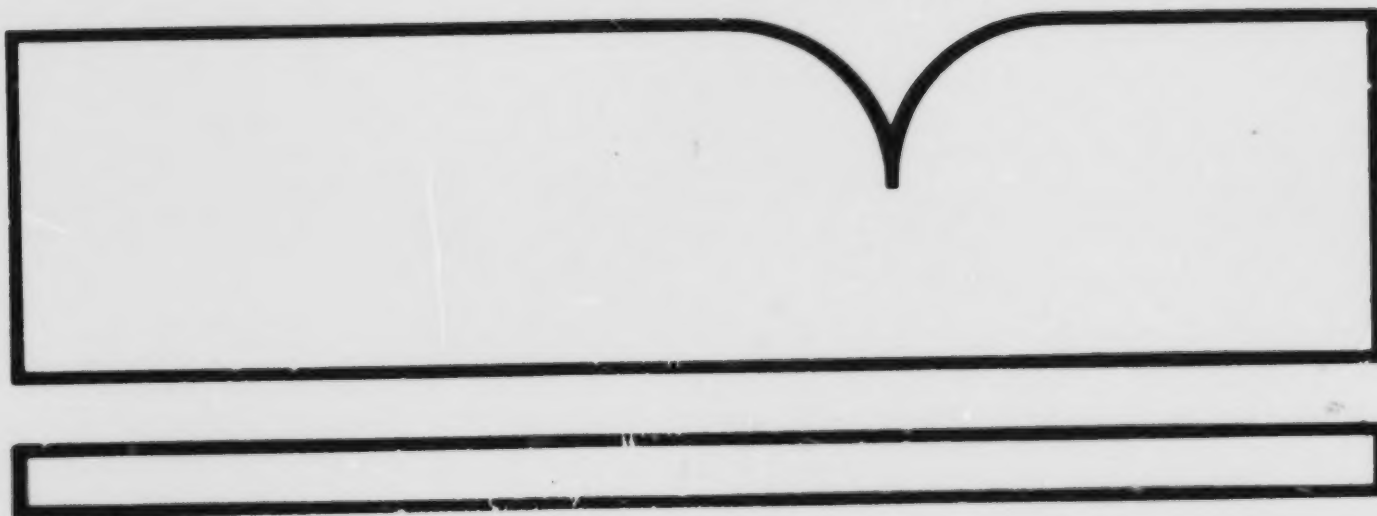
Limestone Scrubber Slurry
Automatic Control Systems

Cincinnati Univ., OH

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LIMESTONE SCRUBBER SLURRY AUTOMATIC CONTROL SYSTEMS

by

Patrick H. Garrett
University of Cincinnati
Cincinnati, Ohio 45221

R805758

D. Bruce Harris
Project Officer
Process Measurements Branch

INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
RESEARCH TRIANGLE PARK, N.C. 27711

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ABSTRACT

The automation of industrial processes is extensive and applied to the purposes of savings of feedstock and increasing the quality of controlled variables. In the case of environmental processes, consistent reduction in the variance of controlled variables with large throughput processes is of particular interest. Accordingly, this report utilizes current understanding of limestone scrubbers for flue gas desulfurization to develop an effort into the optimization of automatic control for the recirculating slurry processes. To this end the acknowledged methods of mathematical modeling, computer simulation, and experimental proofing are applied to the design of slurry limestone addition, slurry density, absorber liquid-to-gas ratio, and slurry buffer addition control systems.

Three automatic control methods are developed, tested, and analyzed for dense limestone feedrate to the recirculating slurry: (1) stoichiometric control based upon material balance considerations, but is compromised due to the lack of slurry reaction measurements; (2) pH control which is geared to accommodate variations in the slurry reactions, but requires a high reaction gain ($\Delta \text{pH} / \Delta \text{limestone feedrate}$) for stable and responsive control; and (3) stoichiometric assisted pH control which offers additional process disorder reduction than either stoichiometric or pH control separately, but its additional complexity is warranted only under conditions of low slurry reaction gain. Absorber liquid-to-gas ratio control for minimizing scrubber energy requirements while maintaining an SO_2 exit target is designed based on a feedforward

SO₂ removal law and pump-matrix selector. Density control of the slurry solids to maintain optimum conditions for both scaling prevention and solids crystallization is also developed, as well as a ratio addition control system for slurry organic acid buffer additives that enhance SO₂ removal mass transfer reactions at the liquid-to-gas interface.

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SECTION 1

INTRODUCTION

Early attempts at the control of limestone scrubbers involved combinations of liquid-to-gas ratio, limestone feedrate, and recirculating solids which would maximize SO_2 removal while minimizing the risks of internal scaling. However, an incomplete understanding of observed effects conspired to prevent achievement of desired goals. More recently the characterization and operating experience with these processes has reached a level of maturity to better support a productive effort into the optimization of their control.

The application of automatic control to environmental processes such as limestone-alkali wet scrubbers is especially promising because of its potential for increased reliability, economy of operation, and consistent reduction in the variance of controlled variables with varying operating conditions and operator proficiency. Implementation of solutions to the major obstacles to limestone scrubbing, the latter including internal scaling and insufficient SO_2 removal, in practice can be achieved only through the disorder reducing capabilities of process automation.

The effectiveness of limestone scrubbing is primarily dependent upon the ability to accommodate and regulate the manifold factors associated with the recirculating slurry. This problem is all the more a challenge because of the limited number of useful slurry process measurements and actuator inputs available, and the economic

considerations associated with the energy and feedstock requirements of large scale scrubbers. These consequences prompted the utilization of the methods of process modeling, computer simulation, and control system experimental testing for development, design, and the detailed analysis of the recirculating slurry control systems presented in this report. Effort is focused upon understanding the slurry processes, and then mating optimum control methods to the specific controlled variables of limestone feedrate, slurry density, absorber liquid-to-gas ratio, and slurry buffer additives which can accommodate their respective slurry factors and minimized process disorder.

SECTION 2

CONCLUSIONS

Effective control of the recirculating slurry is essential for successful operation of limestone scrubbers at the reliability, SO_2 removal level, and efficiency of interest. Three control methods have been analyzed and specified for the addition of dense limestone to the slurry, which constitutes the most difficult scrubber control action to realize effectively because of competing slurry factors.

Stoichiometric control of limestone addition is widespread in application, unconditionally stable, and meaningfully based on slurry material balance measurements. A recommended design for stoichiometric control is detailed in this report. However, analysis shows this control method more accurately to be dead reckoning based on feedforward measurements, and lacking in its visibility of influential slurry factors which can result in significant errors in the limestone feedrate. Therefore, its utility as the sole determiner of limestone addition is compromised and not recommended.

PH control of limestone feedrate is geared to accommodate slurry factors which are invisible to stoichiometric control, primarily because slurry pH measurement represents an agglomerate of the slurry reactions. However, the implementation of this control method requires a rigorous design for stable and responsive performance, since the process intervenes between the controller measurement and limestone feed. Performance is improved with an anti-windup batching controller

which operates to prevent further integration of the error signal when the output actuation signal has limit cycled. A nonlinear pH controller is not required for limestone reagent because of the highly buffered dissolution and limited range of neutralization reaction pH.

A narrow range 4.5 to 6.5 pH probe was found to increase sensitivity and control responsiveness, and cascade flow control was used to compensate for disturbances in the limestone feed system. Responsive pH control is achieved for pH set points (< 5.8) that result in a high slurry reaction gain ($\Delta \text{pH} / \Delta \text{limestone feedrate}$). Consequently, this is the recommended limestone feedrate design if conditions of high reaction gain are maintained during scrubber operation.

The combination of stoichiometric assisted pH control utilizes the most independent process measurements to determine the limestone feedrate, and is optimum in the sense that further reduction in the slurry process disorder is available than either method is capable of individually. The deadband pH and integrating stoichiometric ratio control additions necessary to realize improved performance result in increased system complexity, which is warranted only under conditions of low slurry reaction gain.

Density control of recirculated slurry solids, essential to scrubber scaling prevention and solids crystallization, is a straightforward process readily amenable to automatic control. Similarly, the addition of organic acid buffer additives to the limestone slurry to boost SO_2 removal efficiency was found to require an uncomplicated ratio control mechanization referenced to the limestone feedrate. Recommended designs are provided for both density and buffer addition control systems.

An analysis of limestone slurry liquid-to-gas ratio control for minimum scrubber energy requirements, while simultaneously maintaining an SO_2 exit target, resulted in the design of a feedforward control system. This system uses inlet SO_2 and flue gas volumetric flow measurements with a removal efficiency law and slurry pump selector logic for continuously determining the minimum required L/G ratio. The addition of an outlet SO_2 measurement to compensate the feedforward L/G determination for errors in the exit target was found to produce control system instability under all conditions and therefore is impractical. These results are summarized in Table 1.

Element	Modeled	Simulated	Tested	Highlight
pH control of limestone feedrate	X	X	X	Recommended control method, but requires high reaction gain ($\Delta \text{pH} / \Delta \text{limestone feedrate}$).
Stoichiometric limestone feedrate control	X	X	X	Rapid control response to inlet changes, but errors due to no slurry measurements.
Stoichiometric assisted pH control	X	X	X	Maximum slurry disorder reduction, but complexity warranted only for low reaction gain.
Absorber liquid-to-gas ratio control	X	X		Energy savings based on feedforward control using inlet SO_2 . Unstable using outlet SO_2 .
Limestone slurry density control	X	X		Maintains optimum condition for solids crystallization and scaling prevention.
Slurry buffer additives control	X			Ratio control determined from percent of limestone feedrate.

Table 1. Limestone Scrubber Slurry Automation Research Milestone Chart

SECTION 3

RECIRCULATING SLURRY MATHEMATICAL MODEL

Successful implementation of process automation is dependent on an accurate mathematical model of the system to be controlled. A mathematical model of the recirculating limestone slurry dissolution and neutralization reactions was derived and then validated by means of a regression fit to the Shawnee TCA limestone scrubber depletion runs, obtained from Bechtel¹ and shown in Figure 1. This model defines the absorber-hold tank recirculation path in terms of the three elements of Table 3 and Figure 2. Coefficient values for this model are based upon a typical 10 minute slurry holding tank residence time t_d , and the average slurry reaction gain k_p ($\Delta \text{pH} / \Delta \text{limestone feedrate}$) of 0.18 also obtained from Bechtel's factorial run data² tabulated in Table 2. Note that this reaction gain moves inversely to slurry pH.

¹EPA Alkali Scrubbing Test Facility: Advanced Program Third Progress Report, EPA-600/7-77-105, September, 1977.

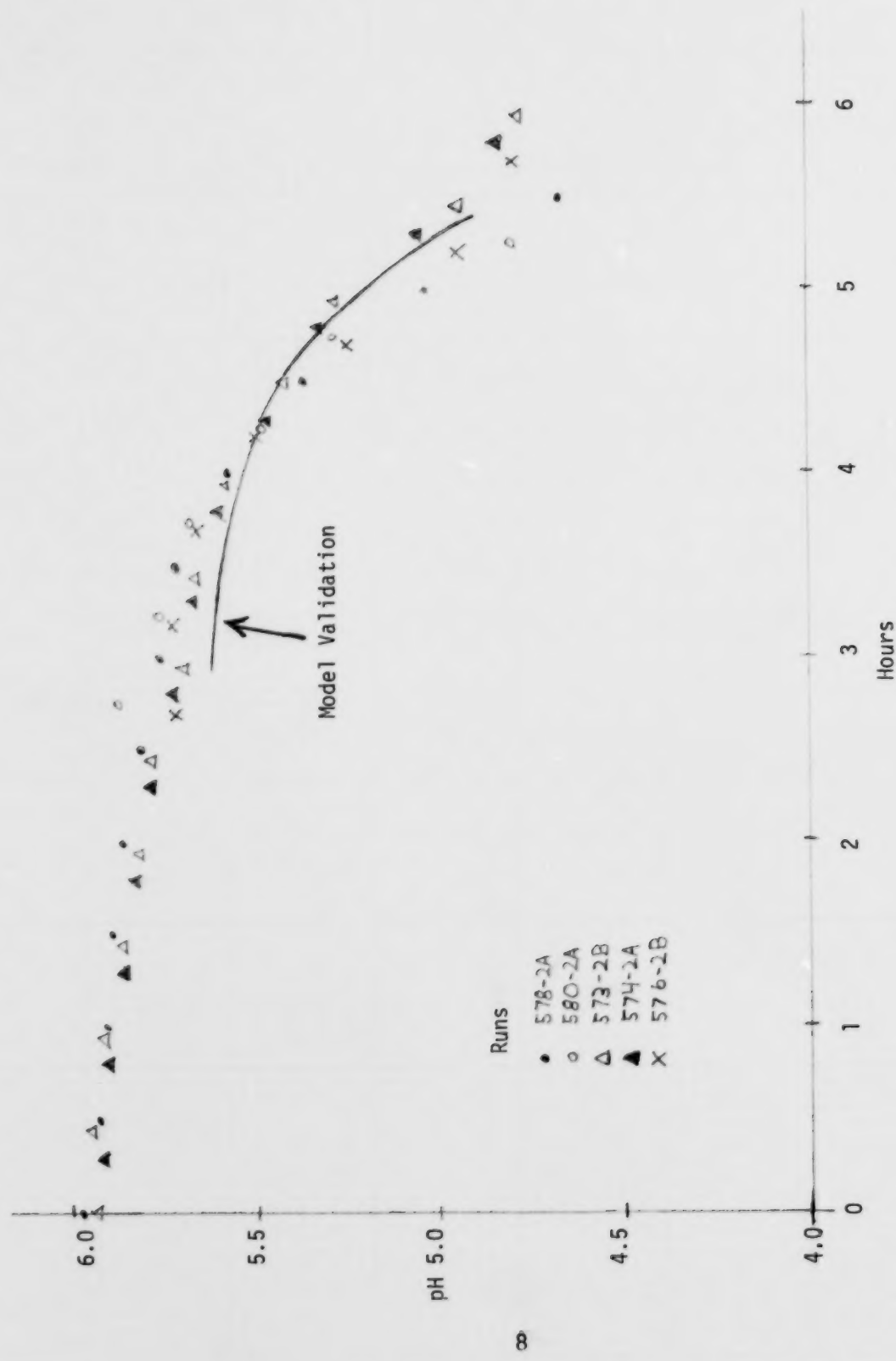


Figure 1. Limestone Slurry Depletion Runs

ΔpH	Δgpm	k_p
4.7 - 5.0	0.8	0.375
5.0 - 5.3	1.1	0.273
5.3 - 5.5	1.6	0.125
5.5 - 5.7	1.6	0.125
5.7 - 5.85	1.4	0.107
5.85- 6.0	1.5	0.100

Table 2. Slurry Reaction Gain

$$\begin{aligned}
 \text{limestone dissolution} &= \frac{k_1}{1 - k_2 \exp(-t_d S)} \quad (1) \\
 &= k_1 + k_1 \cdot k_2 + k_1 \cdot k_2^2 + \dots + k_1 \cdot k_2^{n-1} \\
 &= \frac{k_1 - k_1 \cdot k_2^n}{1 - k_2}
 \end{aligned}$$

The kinetics of limestone dissolution are difficult to model because of the slurry recirculation deadtime characteristic. Deadtime is mathematically represented by the negative exponential in the dissolution function of Table 3, where the coefficients k_1 and k_2 determine the fractions of dissolved to recirculated limestone, respectively. It is notable that this dissolution function was determined to fit the geometric progression identity of equation (1), where $k_1 + k_2 = 1$. These coefficients were determined to be $k_1 = 0.037$ and $k_2 = 0.963$ in order to satisfy the regression fit to the depletion run data, and were verified from the composite response of this model represented by equation (3).

The neutralization time constant for a back-mixed hold tank is essentially the tank volume to flow rate which is equal to $t_d = V/F$, where the reaction rate is assumed to extend over the duration of this residence time. This first-order time constant becomes a pure integrator with the application of Mason's formula to the neutralization function in Table 3, and is due to its position in the slurry recirculation circuit shown in the model of Figure 2. This integrator term provides the single frequency-determining element in the limestone slurry model, which has a -20dB/decade response, and provides an increasing reaction gain for a decreasing frequency of limestone feedrate change (produced by a decreasing frequency of load-change disturbances).

An interesting characteristic of limestone dissolution is the resonance phenomenon illustrated by Figure 3 which arises from the distributed nature of the scrubber. The resonant period is $2\pi/t_d =$

0.628 radian/minute for a 10 minute residence time, and exhibits a peak-to-peak amplitude of 34dB over any 1 radian/minute frequency span for the k_1 and k_2 coefficients determined from the depletion runs of Figure 1. The slope of the amplitude response is a function of the neutralization integrator term, and crosses the 0dB axis at the frequency of 0.0063 radian/minute with 33 degrees of phase margin above an unstable -180 degrees. -90 degrees of initial phase shift at 0 radians/minute is provided by this integrator term, with a superimposed 150 degree peak-to-peak phase variation attributed to the dissolution function. The limestone dissolution and neutralization transfer function is therefore found to be unconditionally stable prior to being connected to a closed-loop control system, but does have a critical frequency of 0.016 radian/minute where the phase margin decreases to 15 degrees. Consequently, operation of the scrubber with load disturbances occurring at intervals of about 1 hour can result in an oscillatory response, and may explain some instances of anomalous scrubber behavior. This process response also has an analogy to that of a crossflow heat exchanger.²

ELEMENT	CHARACTERISTIC	FUNCTION	APPROXIMATION
Dissolution $G_D(S)$	Geometric Progression	$\frac{k_1}{1 - k_2 \exp(-t_d S)}$	$\frac{(0.037) - (0/037)(0.063)^n}{1 - 0.963}$
Reaction Gain K_p	Quasi Linear	$\frac{\Delta pH}{\Delta \text{feedrate}}$	$0.18 \left(\frac{\text{pH}}{\text{liters/min}} \right)$
Neutralization $G_N(S)$	Integrator	$\frac{1/t_d S + 1}{1 - 1/t_d S + 1}$	$\frac{1}{10S} \text{ (min}^{-1}\text{)}$

Table 3. Dissolution and Neutralization Model Elements

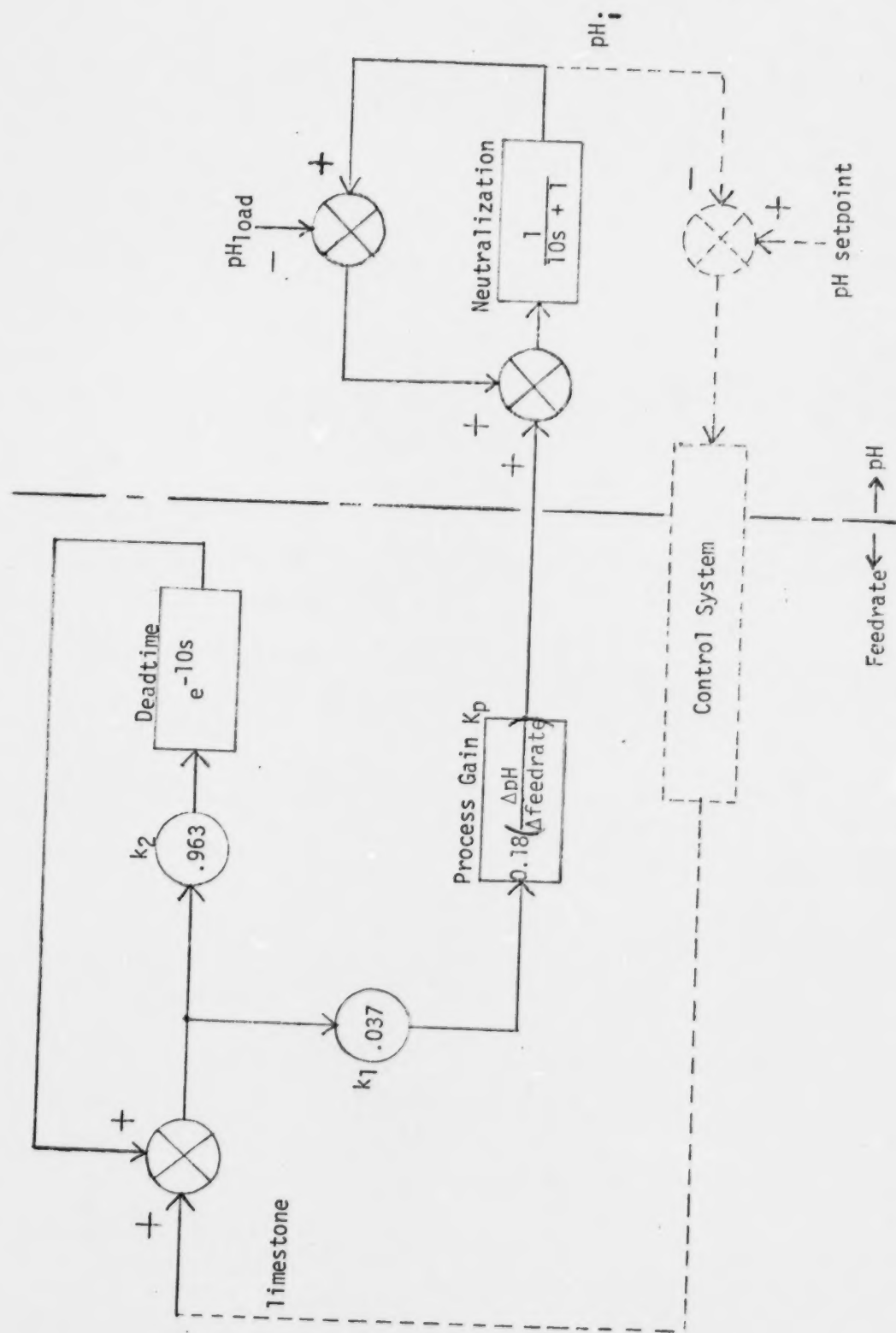


Figure 2. Slurry Dissolution And Neutralization Model

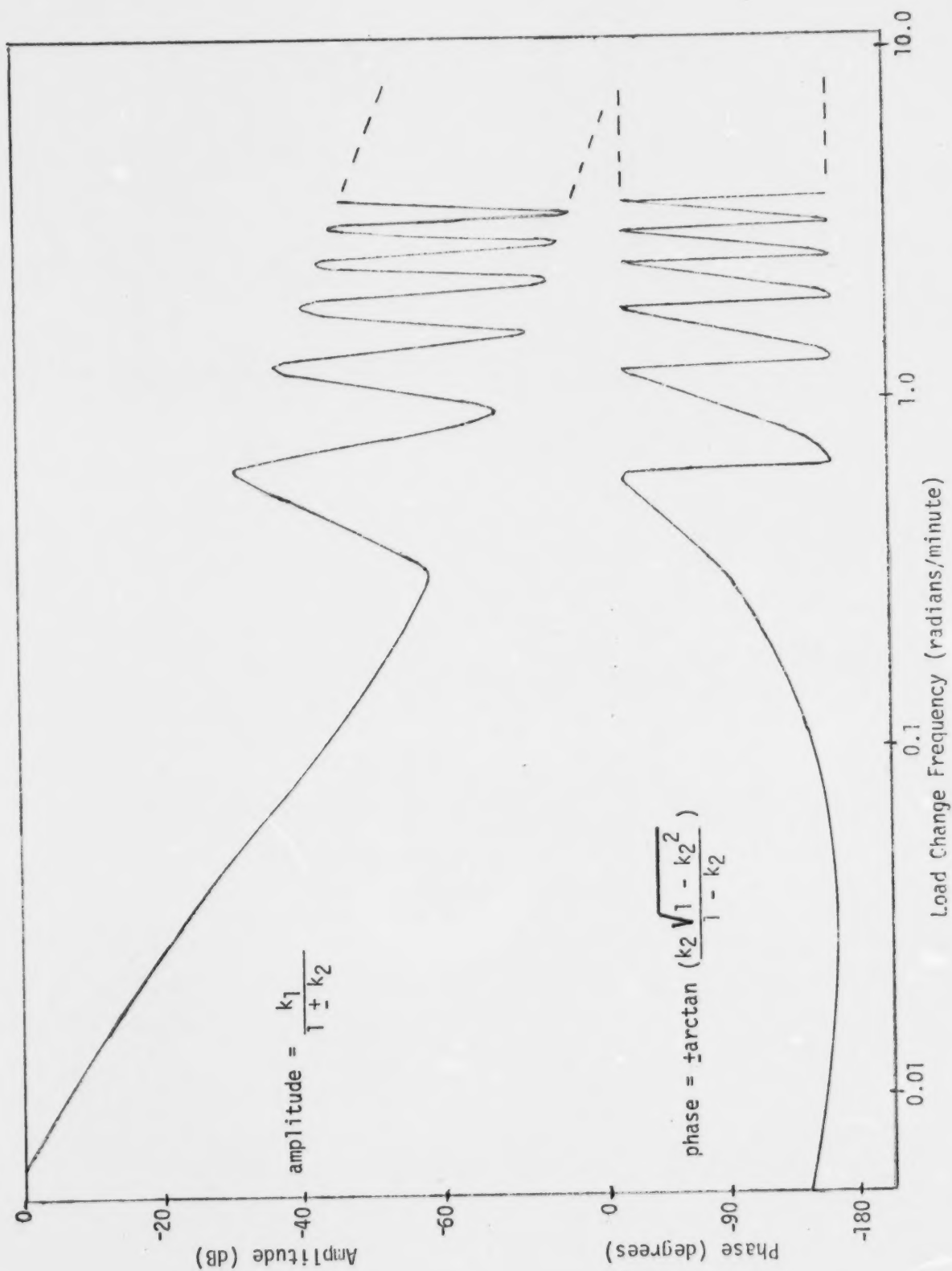


Figure 3. Scrubber Amplitude And Phase Response

Additional analytical insight into the controllability of this process is available by application of the Final-Value Theorem to the closed-loop transfer function, including the addition of a proportional plus integral (PI) process controller to the model transfer function of equation (2) included in equation (4). Results for step and ramp type load changes are described by equations (5) and (6) with zero steady-state error for typical step changes in scrubber load, and a finite error for less frequent ramp changes in load that diminishes with increasing controller gain K . Then term t_r represents the controller integral time that provides an integrator term $1/t_r S$ which combines with the neutralization integrator term $1/t_d S$ to form the second-order system described by equation (4). A second-order system is required in order to track load disturbances. However, its stability is dependent upon a sufficiently large integral time t_r so that its phase-shift contribution does not result in 360 degrees around the closed loop or in overall gain greater than one. These conditions are met through appropriate controller tuning adjustments presented in Section 8.

²Harriott, P. Process Control, pp. 218, McGraw Hill, c.1964.

$$\text{model transfer function} = \frac{\Delta \text{pH}_i}{\Delta \text{feedrate}} = \frac{k_1 k_p}{t_d s [1 - k_2 \exp(-t_d s)]} \quad (2)$$

$$\begin{aligned} \text{model validation=} \quad \mathcal{L}^{-1} \left[\frac{\Delta \text{feedrate}}{s^2} \cdot \frac{k_1 k_p}{t_d [1 - k_2 \exp(-t_d s)]} \right] \quad (3) \\ \text{depletion} \\ \text{response} \\ = 6.03 \left[1 - \exp \left(-4.5 + \frac{t}{2} \right) \right] \end{aligned}$$

$$\begin{aligned} \text{closed loop} \\ \text{control model} = \frac{\Delta \text{pH}_i}{\Delta \text{pH}_{\text{load}}} = \frac{-\frac{s}{t_d} [1 - k_2 \exp(-t_d s)]}{s^2 [1 - k_2 \exp(-t_d s)]} \quad (4) \\ + \frac{k \cdot k_1 \cdot k_p}{t_r t_d} \cdot (t_r s + 1) \end{aligned}$$

$$\begin{aligned} \lim_{s \rightarrow 0} s \cdot \text{pHi}(s) \Big|_{\text{pH}_{\text{load}}(s) = \frac{1}{s}} = 0 \quad (5) \end{aligned}$$

$$\begin{aligned} \lim_{s \rightarrow 0} s \cdot \text{pHi}(s) \Big|_{\text{pH}_{\text{load}}(s) = \frac{1}{s^2}} = \frac{-\frac{1}{t_d} (1 - k_2)}{k \cdot k_1 \cdot k_p / t_r t_d} \quad (6) \end{aligned}$$

SECTION 4

SIMULATION OF REAGENT CONTROL METHODS

An analog computer was utilized for simulation studies of the slurry model and was instrumental in providing insight into the process dynamics and control requirements for limestone scrubbers. Figure 4 shows a composite of the process and controller mechanizations simulated. A PI controller option that provided additional control system stability was an antiwindup batching feature, which disconnects the error signal to the integrator for output actuation commands of $<0\%>100\%$. This prevents the controller from deleteriously integrating the error signal further when its actuation capability has limit cycled. The control system was simulated with step changes in load of ± 0.2 pH unit, which correspond to ± 1000 ppm changes in SO_2 . For all runs the slurry pH setpoint was maintained at 5.5 with conversion of the incremental slurry model into an absolute model for computer simulation purposes.

The pH control simulation results are summarized by Figure 5, and show the shortest time for recovery to the operating setpoint following load changes is provided by the combination of feedforward and feedback control of Run (a). This reflects the ability of feedforward control to respond to inlet changes in SO_2 prior to influencing the slurry, with pH feedback control geared to solve for slurry unknowns such as flyash alkali content. The Smith predictor of Run (b) does not provide the anticipated control improvement with its compensatory algorithm, which

utilizes the process deadtime information t_d to make more intelligent corrections over that possible with the delayed process error signal alone. The reason is evident from Figure 4, where the lag elements of both the neutralization integrator $1/t_d s$ and controller integrator $1/t_r s$ unavoidably are in the path of the load change information on its circuit to the Smith predictor input. This unfortunately defeated the purpose of the Smith predictor; i.e., to transform the visibility of the process to the controller to appear as if the process has no time delay.

Runs (c) and (d) describe the process response with straight pH feedback, which was determined to be capable of accurate and well-behaved control. Based on the criteria of deviation of the controlled variable from setpoint following a load disturbance and recovery time to setpoint, the controller antiwindup batching feature was found to offer advantage over a conventional proportional plus integral controller. Although load increases require approximately the same recovery time with and without the antiwindup batching controller feature, the more troublesome load decreases representing excess limestone entrainment in the slurry recovered to setpoint in a notably shorter interval with it. Coordination of Figure 5 and the slurry process model response of Figure 3 is possible from observation that the identical 220 minute periods of the scrubber pH response immediately following a load decrease in Figure 5, are equivalent to 0.03 radian/minute for a pH variation of -16dB (0.125) in Figure 3 (0.7 pH unit positive excursion from a pH setpoint of 5.5).

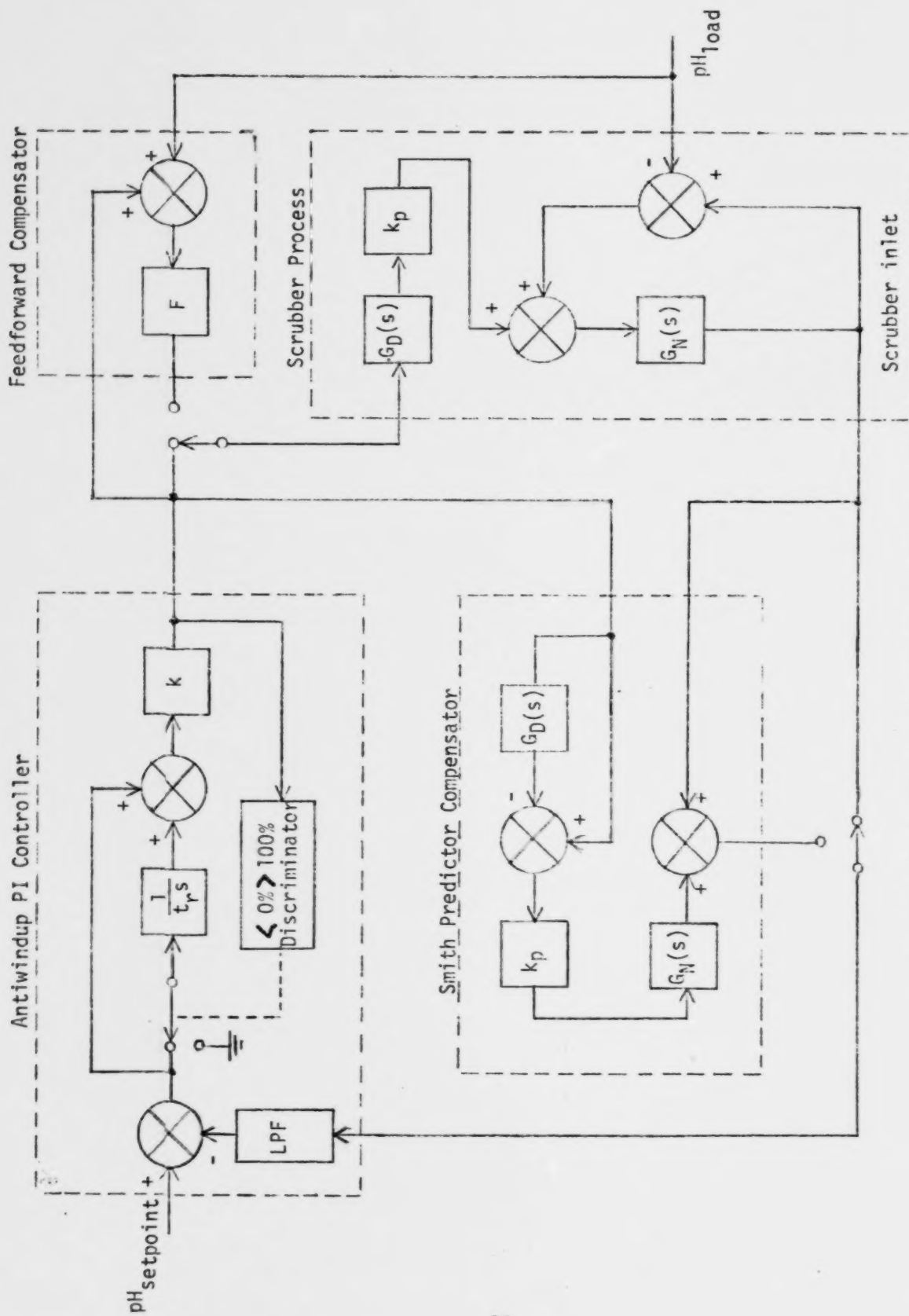


Figure 4. Composite Of Simulated Control Methods

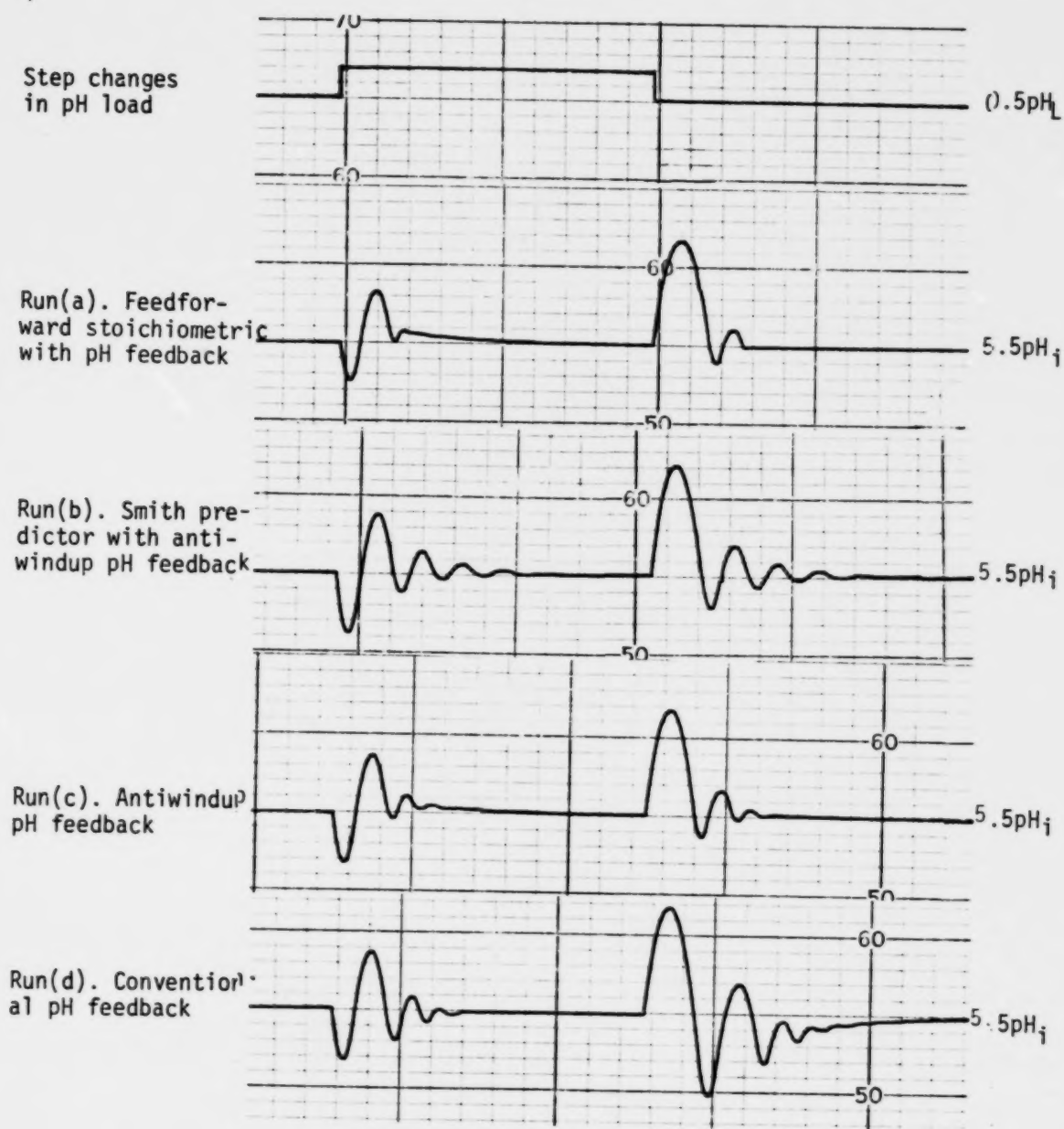


Figure 5. Reagent Control Simulation Runs

Legend: 100 minutes/horizontal division
0.1pH/vertical division

For these simulation exercises a controller gain k of 100 (1% proportional band) and an integral term t_r of 200 minutes were utilized, where the latter was determined to be the natural oscillation period of the closed-loop system. These values provided responsiveness with stability for simulation purposes, but will be superceded by actual tuning values on a scrubber system. Table 4 provides a summary of simulation parameters and results. All of these simulation exercises utilize the scrubber model values presented in Table 3 of Section 3.

A properly designed analog computer simulation requires both time and magnitude scaling of the equations representing the system to be evaluated, and their manipulation into a form that can be directly implemented by available computer functions. The scrubber model deadtime and residence time of 10 minutes simplified time scaling and resulted in a timing scale factor of 1/600, whereby 1 second of computer run time corresponds to 10 minutes of the process. The largest element mechanized by the computer was the deadtime function e^{-10S} , implemented with a fourth-order Pade' approximation shown by the six amplifier cascade (9 through 14) in Figure 6. Magnitude scaling was used to limit the computer function outputs to 10 volts full scale on the EAI TR-20 computer. The simulation variables and normalization coefficients are listed in Table 5.

The simulated controller output is limited between 0 and 100 per cent by the symmetrical diode limiter, and the antiwindup controller feature was implemented with a differential comparator connected to the controller output driving a relay to switch the integrator input to

ground during controller output limiting conditions. The four models of the scrubber slurry control methods simulated were designated Models A through D. Model A was pH feedback control with a conventional PI controller. Model B represents the same method with the addition of the controller antiwindup feature. Model C was the pH feedback augmented feedforward control system with the PI antiwindup controller. Model D represents pH feedback control with the antiwindup feature and a Smith linear predictor compensator. An austere first-order deadtime approximation employing amplifier 18 was used in the linear predictor model because of computer hardware limitations.

RUN	METHOD	CONTROLLER	pH LOAD	RECOVERY
(a)	Feedforward plus pH feedback	Antiwindup PI $k = 100$ $t_r = 200$ min. $F = 10$	5.5 ± 0.2	300 minutes
(b)	pH feedback with Smith predictor	Antiwindup PI $k = 100$ $t_r = 200$ min.	5.5 ± 0.2	600 minutes
(c)	Antiwindup pH feedback	Antiwindup PI $k = 100$ $t_r = 200$ min.	5.5 ± 0.2	350 minutes
(d)	Conventional pH feedback	Conventional PI $k = 100$ $t_r = 200$ min.	5.5 ± 0.2	900 minutes

Table 4. Summary of Simulated Reagent Control Methods

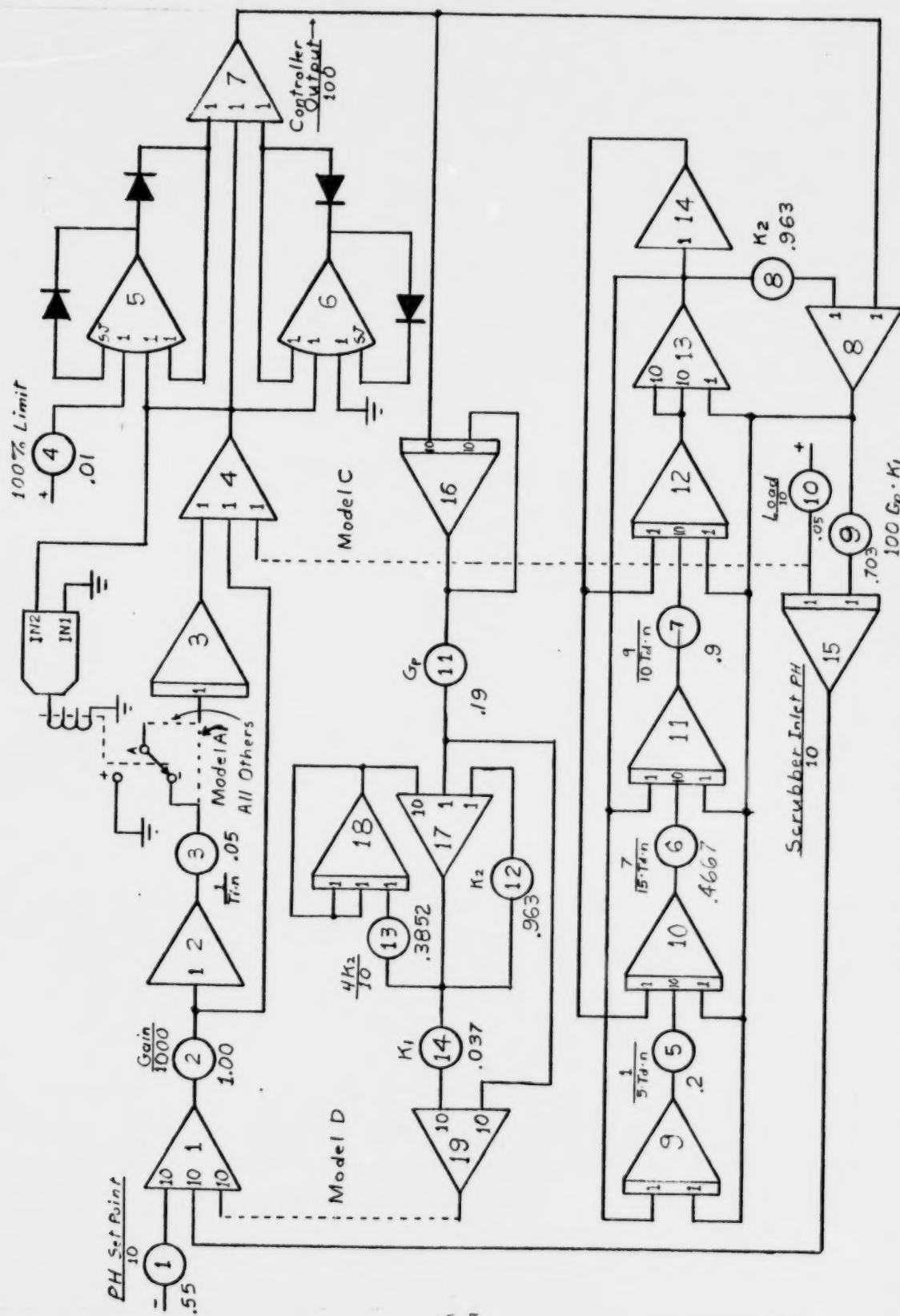


Figure 6. Analog Computer Simulation Schematic

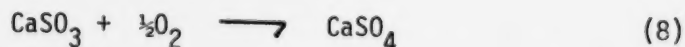
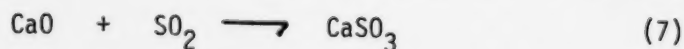
Variable	Normalized	Multiplier	Description
pH_S	x_{phs}	10	Scrubber pH setpoint
pH_L	x_{phl}	10	Scrubber pH load
pH_I	x_{phi}	10	Scrubber inlet pH
n	n	1/600	Time scaling constant
k_1	k_1	0.037	Limestone dissolution constant
k_2	k_2	0.963	Deadtime dissolution constant
t_r	t_r	10 min	Controller integral time
t_c	t_c	10 min	Neutralization time constant
G_p	x_{gp}	100	Scrubber process gain
A_c	x_{ac}	1000	Antiwindup PI output
A	x_a	10	Smith predictor neutralization
B	x_b	100	Smith pH dissolution
C	x_c	1000	PI controller output
D	x_d	100	Smith deadtime
E	x_e	1	pH error signal
F	x_f	10	Smith predictor output
L	x_l	1000	Limestone addition
P	x_p	10	Scrubber pH dissolution

Table 5. Variables and Normalization Multipliers

SECTION 5

STOICHIOMETRIC CONTROL SYSTEM DESIGN

Absorption of SO_2 by a limestone slurry in a wet scrubbing process is normally followed by subsequent oxidation of the calcium sulfite to calcium sulfate described by equations (7) and (8). If the scrubber physical chemistry is straightforward, the stoichiometric ratio of limestone to sulfur corresponding to the desired sulfur removal can be calculated and the dense limestone addition provided under stoichiometric control. In practice, limestone is added at a rate proportional to the inlet mass SO_2 rate coordinated with a setpoint equated to the desired stoichiometric ratio (SR), as shown in Figure 7. It should be noted that this stoichiometric ratio setpoint is, in fact, only a feedforward gain adjustment which is prevented from agreeing with the true slurry stoichiometry by the primary factors of flyash alkaline content, limestone quality and solubility, and chloride content of the slurry. Nevertheless, feedforward control is useful for limestone feedrate because of its ability to correct for SO_2 load disturbances to the sluggish buffered recirculating slurry. However, its utility as the sole determiner of limestone addition is compromised as described in subsequent sections.



Feedforward limestone addition systems referenced to stoichiometric ratio control are based upon the material balance of moles of SO_2 absorbed to moles of Ca added. Real-time computation of these parameters

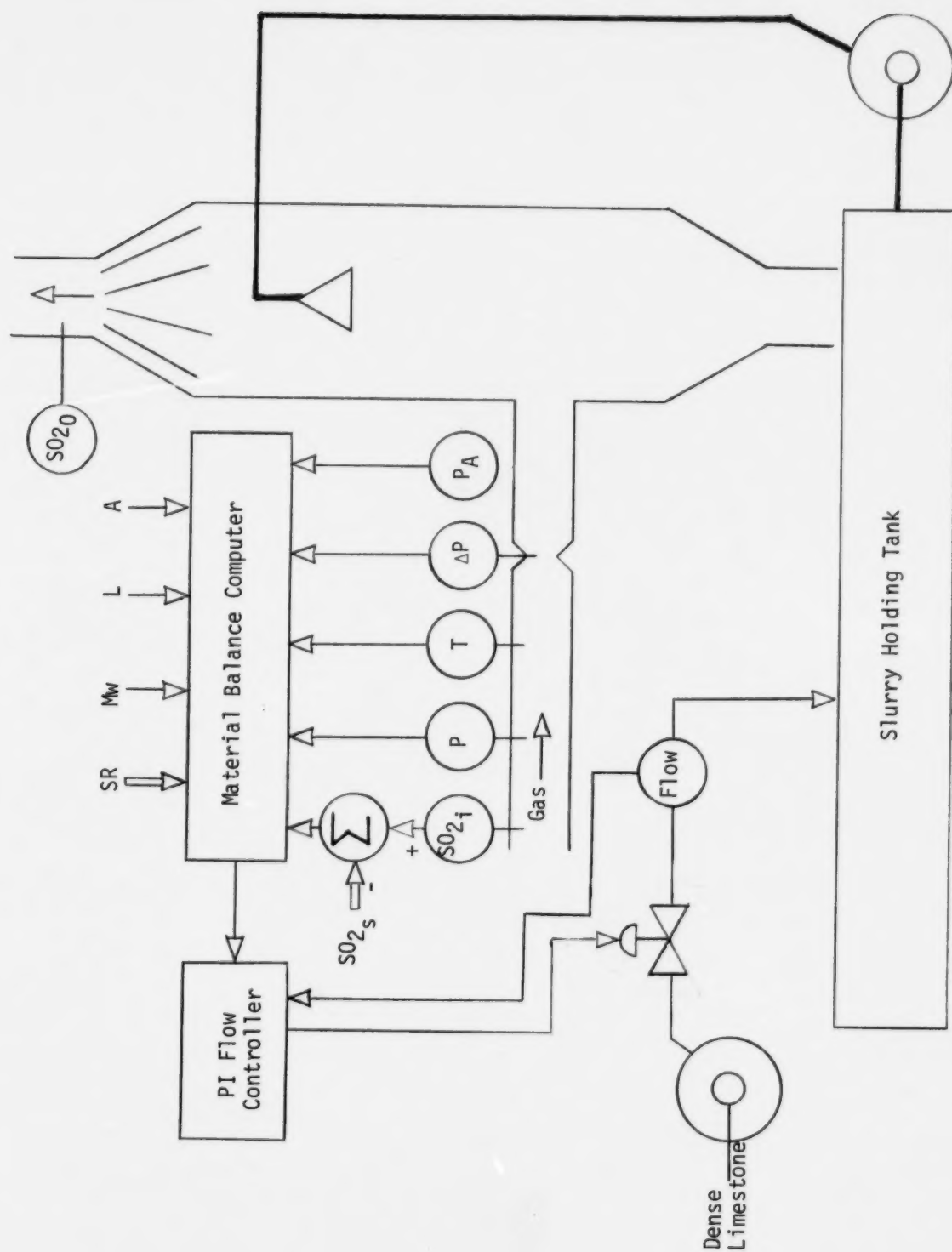


Figure 7. Feedforward Stoichiometric Limestone Addition Control

can be provided from the scrubber flue gas inlet duct by means of a dedicated computation unit using temperature, pressure, and SO_2 analyzer measurements. A compact analog computer which provides this data to the requirements of Method 2 of the U.S. Environmental Protection Agency is shown in Figure 8, Table 6, and equations (9) through (11). The SO_2 inlet concentration has subtracted from it the allowable scrubber SO_2 exit concentration setpoint yielding a ΔSO_2 input to the mass flow computer. Calculation of the slurry limestone addition volumetric flow rate F is then based upon neutralization of the required SO_2 at the anticipated stoichiometric ratio.

The circuit of Figure 8 is implemented with seven standard multi-function modules available from several manufacturers, where each implement the equation $Y.(Z/X)$. The 16 mathematical operations calculate a limestone addition signal within an error of percent, and may be scaled for scrubbers sized to boilers of up to 100 MW capacity and inlet SO_2 concentrations to 4000 ppm. The ΔP signal is provided from a scrubber inletduct venturi, and the only manual operation required is a periodic wet chemistry orsat determination of the total flue gas molecular weight M_w .

$$\begin{aligned}
 Q &= A.k. \frac{\Delta P.T}{P.M_w} \cdot \frac{P.T_{std}}{P_{A.T}} \quad (\text{M}^3/\text{Sec}) \quad (9) \\
 &= M^2 \frac{(\text{g/gmole})(\text{mm}^3/\text{g})M^2}{(\text{oK})(\text{mmHg})(\text{Sec})^2} \frac{(\text{mmHg})(\text{oK})}{(\text{mmHg})(\text{moles})(\text{mmHg})(\text{oK})} \\
 &= \frac{M^3}{\text{Sec}}
 \end{aligned}$$

$$SO_2 = SO_{2_i} - SO_{2_s} \text{ (ppm)} \quad (10)$$

$$M = Q \cdot SO_2 \cdot M_{wSO_2} \quad (11)$$

$$= \frac{M^3}{\text{sec}} \cdot \text{ppm}_{SO_2} \cdot \text{moles}_{SO_2}$$

$$= \text{Kg/Sec}$$

$$L = \text{Limestone\% purity} \cdot \text{density} \cdot \text{specific gravity (kg/liter)} \quad (12)$$

$$F = \frac{M \cdot SR \text{ (liters/second)}}{L} \quad (13)$$

With reference to Figure 7, five measured variables and four settable coefficients are input to the feedforward stoichiometric material balance computer. The output limestone volumetric flow rate F provides the remote setpoint input signal to a proportional plus integral limestone flow controller. This mechanization provides for more accurate limestone addition, by overcoming disturbances in the limestone addition system such as pump cavitation, than the direct application of the flow signal to the valve. Rigorous analysis of this control system shows that it is not truly feedforward because of the existence of the SO_2 exit target setpoint in the material balance equations. True feedforward control requires the substitution of outlet SO_{2_0} for SO_{2_s} to eliminate the error at all times when $SO_{2_0} \neq SO_{2_s}$. This error is representative of the limitations of feedforward control which is completely dependent upon the accuracy to which it models the process. Consequently, this control method is more accurately dead reckoning based on process material balance measurements than feedforward stoichiometric control.

stoichiometric control.

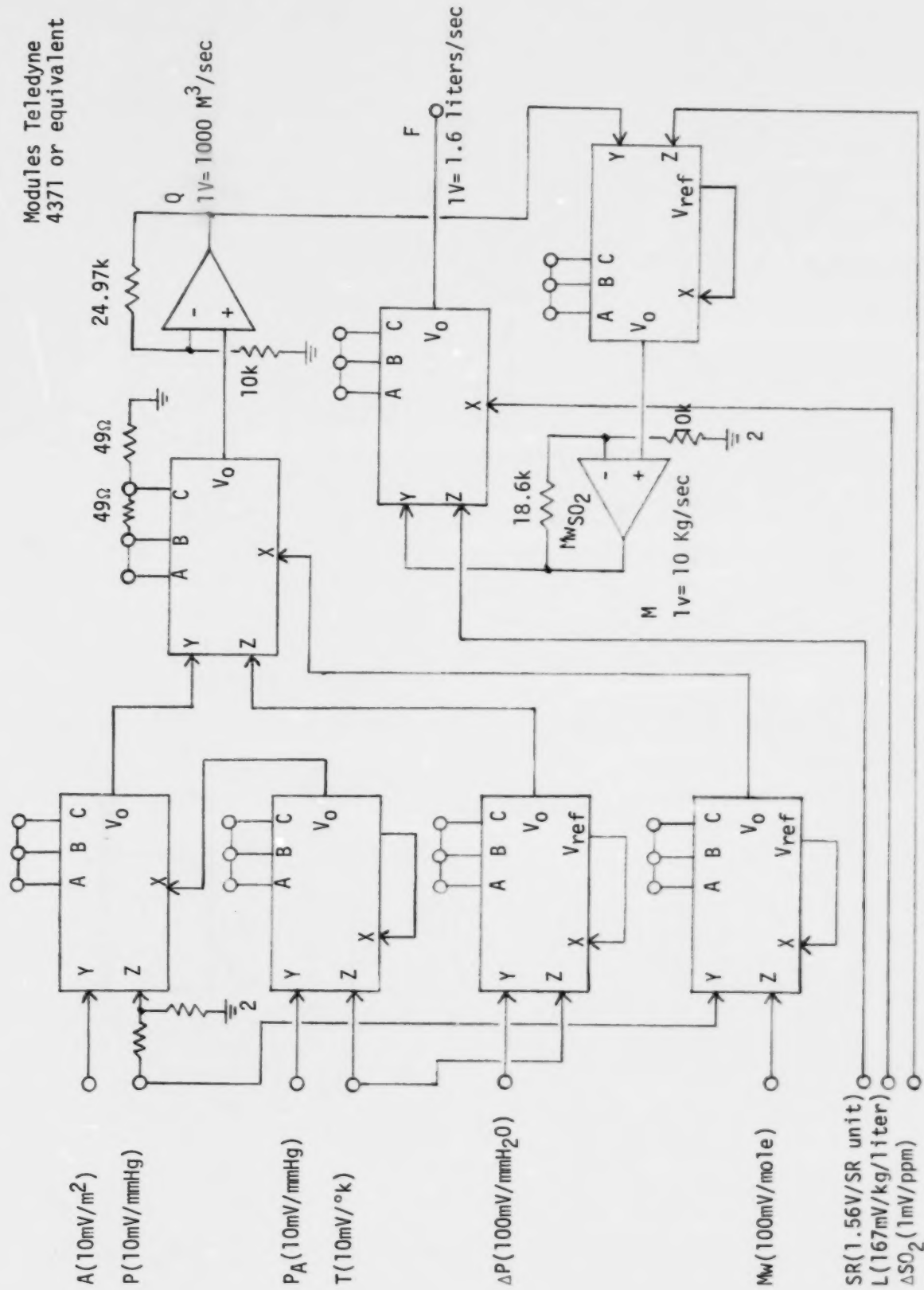


Figure 8. Material Balance Computer Diagram

Parameter	Scaling	Description
A	10mV/m ²	Duct area m ²
P	10mV/mmHg	Duct absolute pressure mmHg
Tstd	0.293 attn	Standard temperature °K
P _A	10mV/mmHg	Absolute atmospheric pressure mmHg
T	10mV/°k	Duct absolute temperature °K
ΔP	100mV/mmH ₂ O	Duct differential pressure mmH ₂ O
M _w	100mV/mole	Flue gas molecular weight moles
ΔSO ₂	1mV/ppm	SO ₂ to be removed
M _w SO ₂	$\frac{64}{22.4}$ gain	Scaled SO ₂ molecular weight moles
L	167mV/kg/liter	Dense limestone Kg/liter
SR	1.56v/SR unit	Stoichiometric ratio $\cdot \frac{\text{mole CaCO}_3(100)}{\text{mole SO}_2(64)}$
k	3.497 gain	$\frac{\text{g/gmole} \cdot \text{mmHgm}^2}{\text{°k} \cdot \text{mmH}_2\text{O sec}^2}$

Table 6. Material Balance Computer Parameters

SECTION 6

PH AND BUFFER CONTROL SYSTEM DESIGN

Feedback control of limestone addition by means of a slurry pH measurement offers the capability of accommodating the scrubber factors listed in Table 7 which feedforward stoichiometric control cannot correct for. However, the control system design is substantially complicated with feedback control because the limestone dissolution and holding tank neutralization reactions are now positioned between the limestone addition and pH measurement as shown in Figure 9. Consequently, a detailed analysis of the scrubber process is required for the determination of a compatible controller which can maintain a required pH setpoint with unconditional stability. The design procedure applied was to utilize a mathematical model which describes the scrubber process, develop a computer simulation of candidate control systems with this model, evaluate the dynamics and stability with likely controller mechanizations under both steady-state and load-disturbance conditions, and proof the final control system design by evaluating it on an actual scrubber.

Reaction gain
Fly ash alkalinity
Calcium sulfite scaling
Limestone variability and solubility
Slurry chloride content

Table 7. pH Visible Slurry Factors

The reagent addition control requirement for limestone scrubbers is essentially that of a batching operation, and notably does not require a nonlinear pH controller because of the highly buffered limestone dissolution and limited range of the neutralization reaction pH. Consequently, a proportional plus integral (PI) controller is appropriate for feedback control because its integral term combines with the integrator function represented by the recirculating slurry to form a second-order system, which is necessary to track process disturbances. In addition, a narrow range 4.5 to 6.5 pH probe is recommended as a practical measure to increase the sensitivity and control responsiveness to the pH measurement.

Major obstacles to the effectiveness of limestone scrubbers have been internal scaling and insufficient SO_2 removal. The ability of pH control to regulate the limestone slurry factors presented in Table 7 results from the use of a slurry pH measurement, which is an indicator of aggregate slurry chemical reactions, with an appropriate control law. Fly ash alkali, limestone alkalinity, and chloride content from coals and makeup water containing HCL all affect the slurry alkali utilization and can be regulated under pH control. This is especially important because of the relationship of alkali utilization to calcium sulfite solubility and the effect of the latter on operating reliability due to internal scaling. For example scaling is normally prevented at a slurry pH of about 5.6 and below, which produces a desirable alkali utilization of 85% and above. However, lower slurry pH values produce reduced SO_2 removal resulting in a narrow range of practical slurry pH operating points about 5.6, which provide a consistent baseline of SO_2 removal generally in the 70% range.

The analytical development of a recirculating slurry mathematical model in Section 3 disclosed that the scrubber absorber-hold tank is an unconditionally stable process with one frequency determining element represented by the integrator term of Table 3. Application of the Final-Value Theorem in equations (5) and (6) to the closed-loop transfer function, which includes the PI controller, also provided zero steady-state error to 1000 ppm step-type load changes in SO_2 . The simulation of this pH control system design represented by Figure 9 indicates, in Section 4, an accurate capability for control of the limestone slurry whose stability is further enhanced with an antiwindup batching option. This PI controller feature operates to prevent further integration of the measurement error signal when the output actuation signal has limit cycled.

Both limestone and organic acid addition are actuated using cascade control, whereby secondary flow controllers compensate for disturbances in the respective feed systems to achieve accurate implementation of the primary controller flow commands. Stable and responsive cascade control requires that the gain of the flow loop (flow controller times reaction gain) be four times or greater than the gain of the primary controller. Equation (14) describes the requirement that the gain of a total closed-loop system cannot exceed unity to realize unconditional stability.

$$\begin{aligned} \text{closed-loop gain} &= (\text{primary gain})(\text{flow gain})(\text{reaction gain}) \\ &\leq 1.0 \end{aligned} \quad (14)$$

With sulfite solubility maintained and scaling prevented with pH control, SO_2 removal can be boosted to the 90% range by the ratio-control addition of organic acid buffer agents such as adipic acid also included in Figure 9. This ratio flow controller adds buffer to the slurry in proportion to $\frac{1}{2}\%$ of the limestone feedrate. In addition, a high alkali utilization factor offers improved controllability of the slurry process because of an increased neutralization process gain ($\Delta \text{pH} / \Delta \text{limestone feedrate}$), and also reduces the amount of undissolved limestone in the sludge. Both controllability and sludge minimization can be further enhanced with forced slurry aeration due, respectively, to CO_2 stripping accelerating limestone dissolution and oxidation of the calcium sulfite to calcium sulfate.

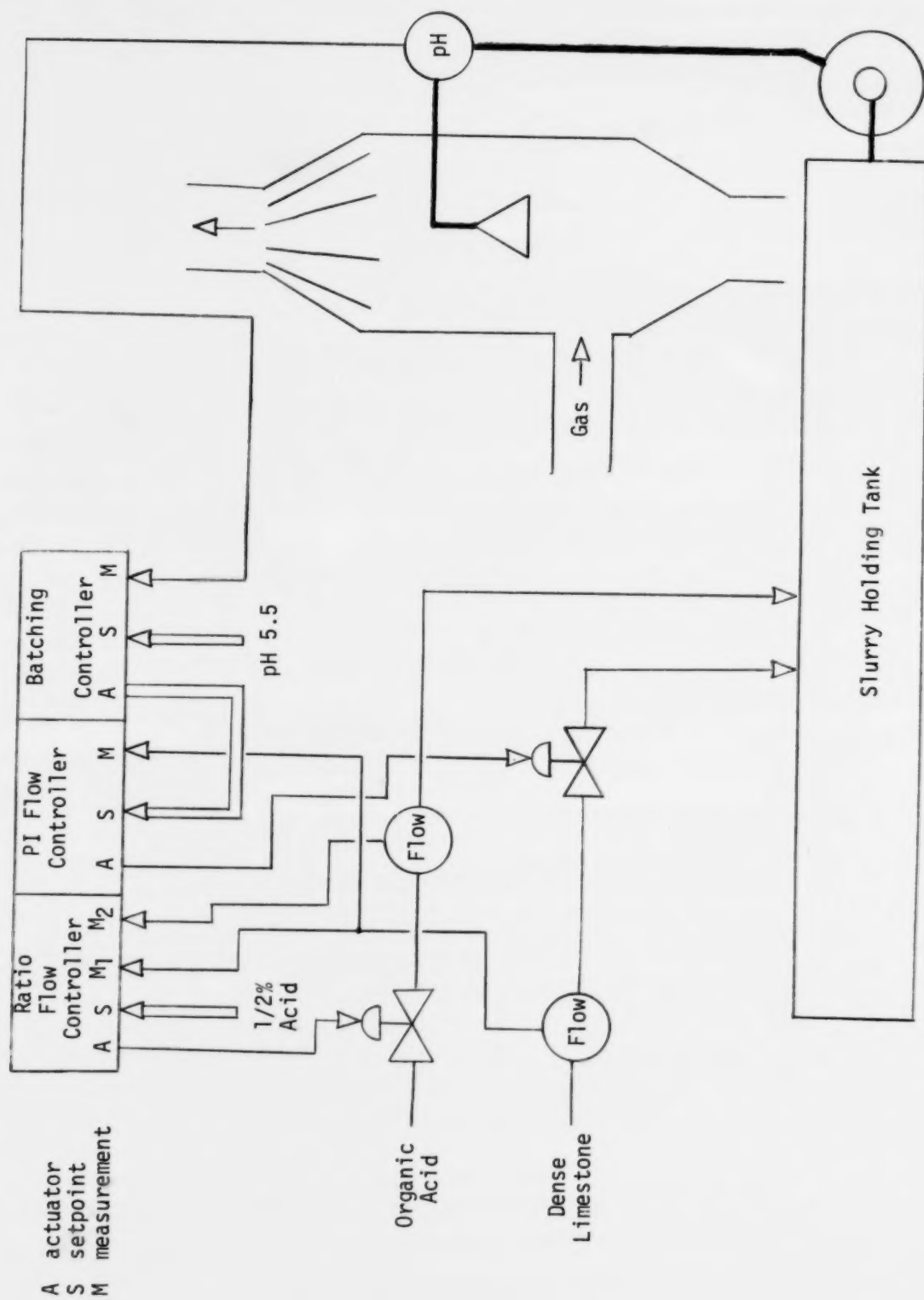


Figure 9. PH Feedback Slurry Control System With Buffer

SECTION 7

STOICHIOMETRIC ASSISTED PH CONTROL SYSTEM DESIGN

The simulation study of pH control with feedforward compensation in Section 4, plus the independent analysis³ of SO₂ removal correlation with pH, both indicate improved regulation of the limestone feedrate is possible when scrubber inlet gas parameters are utilized as well as liquid-phase pH. Application is especially recommended for low slurry reaction gain operation (below 0.25 $\Delta\text{pH}/\Delta\text{gpm}$) to improve controllability of limestone feedrate. Implementation involves combination of the control system designs described by Figures 7 and 9, and capitalizes upon their respective process material balance and slurry pH information to reduce disorder below that possible with stoichiometric or pH control separately.

Figure 10 describes the parallel combination of these two control methods, which is advantageous for stability considerations, with both scaled for the full output capability of the limestone addition system. The significant features of this system are: (1) slurry pH is the primary controlled variable and employs a narrow-range measurement (4.5 - 6.5 pH) to effectively increase control responsiveness without additional controller gain; (2) a computed Ca/SO₂ feedforward material balance prediction using inlet SO₂ and gas rate Q normally determines limestone flowrate; (3) the feedforward gain (stoichiometric ratio SR) is automatically adjusted by an integrating-only controller to accom-

³"Operation of Shawnee TCA with Automatic pH Control," GCA Corporation, August, 1979.

moderate fly ash alkalinity and chlorides in the slurry in response to differences between the setpoint and slurry pH values; (4) the feedback deadband PI controller provides automatic backup pH control in the event of a feedforward system malfunction or major process upsets resulting in slurry pH excursions outside a preselected deadband as illustrated in Figure 11; and (5) cascade control of limestone flow is included to accommodate disturbances in the dense limestone addition system.

To appreciate the logic of this combined control system it must be accepted that the stoichiometric ratio input as defined for material balance limestone feedrate control is actually the feedforward gain setting, that errors in feedforward stoichiometric control are primarily attributable to the slurry factors of Table 7 which pH measurements are sensitive to, and with the proper sense of direction (decreasing stoichiometric ratio with increasing pH) the integrating controller can adaptively trim this feedforward gain to track process anomalies and maintain accurate feedforward control. Figure 11 illustrates this adaptive stoichiometric ratio envelope control relationship. The solid gain curve for the deadband controller has a low value within the deadband zone, and an increasing value for external pH excursions. The stoichiometric ratio feedforward gain is adaptively adjusted along the vertical axis by the integrating controller, following a startup estimate which is preset for the stoichiometric ratio (SR) initial value. An accurate SR gain will coincide with the center of the pH deadband setpoint and zero pH gain.

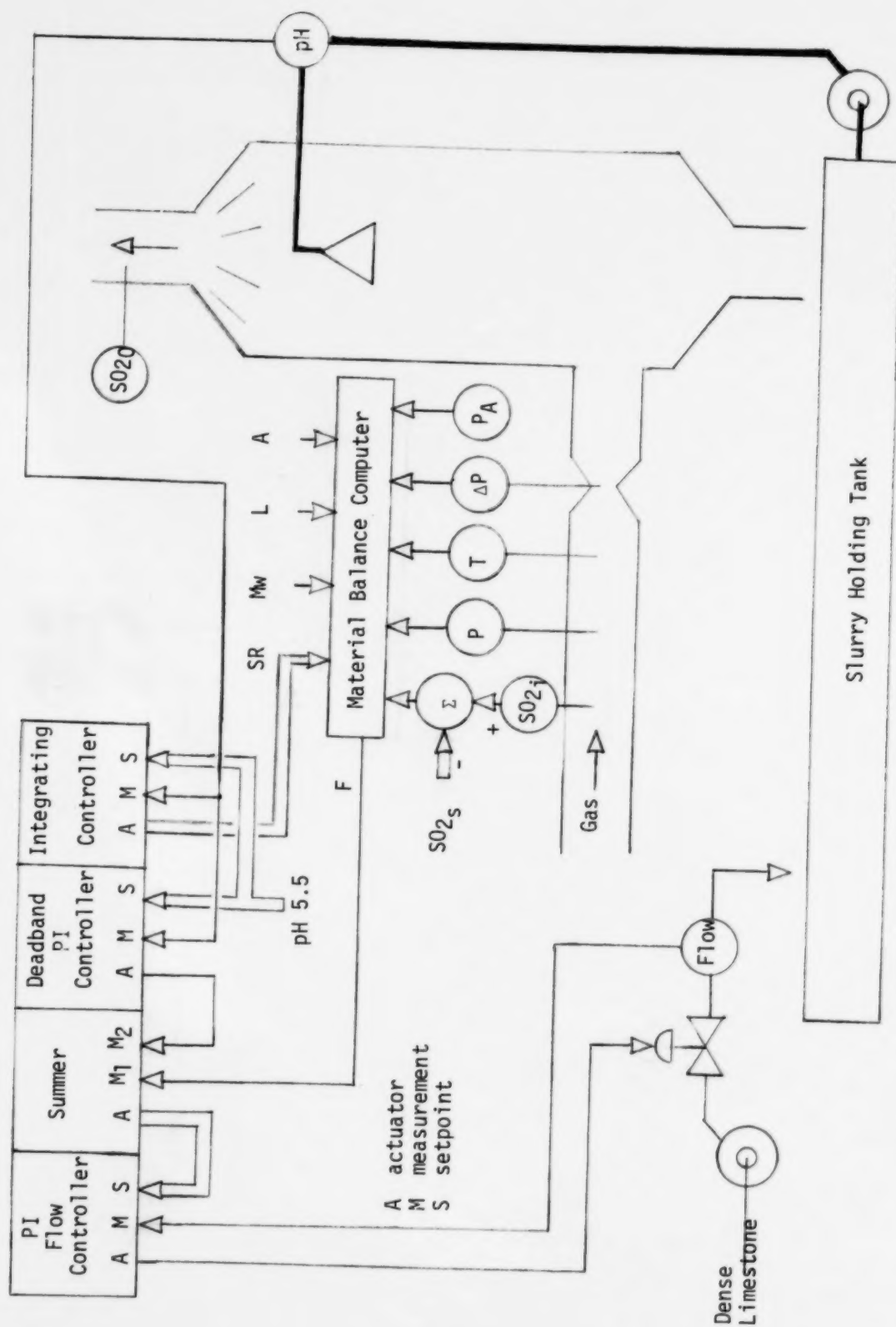


Figure 10. Stoichiometric Assisted pH Control System

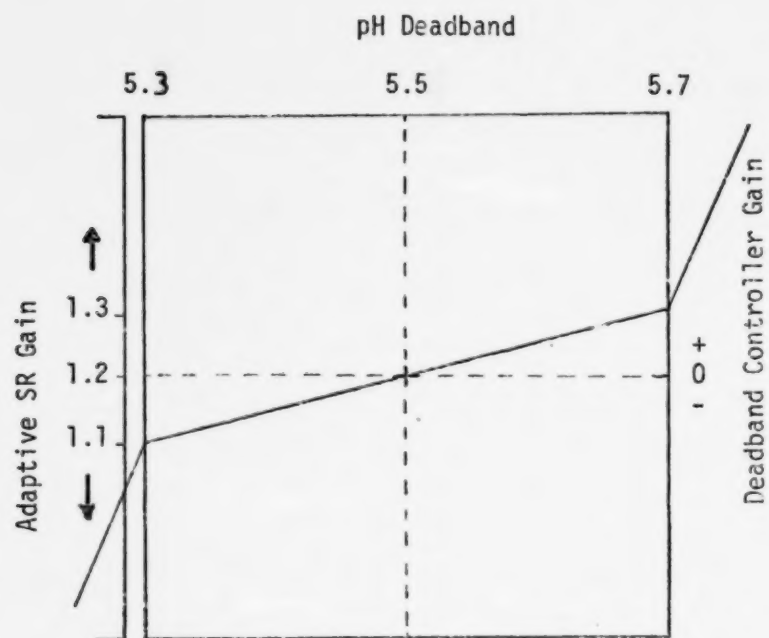


Figure 11. Stoichiometric Assisted pH Control Relationship

To obtain additional insight into the performance of stoichiometric assisted pH control of limestone feedrate, a closed-loop analytical model was derived as presented in Figure 12 and Table 8 utilizing parameters from Section 3 where appropriate. This model is valid for operation within or outside the pH deadband, and is mathematically represented by the transfer function of equation (15) derived in Appendix A. This equation expresses the effect at the absorber inlet on slurry pH_i as a result of SO_2 load changes represented by pH_L .

Application of the Final-Value Theorem to this transfer function with both step-type and ramp-type load changes is described by equations (16) and (17), respectively, and yields zero steady-state error. This result predicts an improvement in control performance and stability in comparison with either stoichiometric or pH control of limestone feedrate used separately. Cascade control is also utilized in this control system design, whereby a secondary flow controller compensates for disturbances in the dense limestone feed system to achieve accurate implementation of flow command signals.

$$pH_i = \frac{-pH_L(s)}{\left[\frac{(tr_4s + 1)k_4}{(1 + k_4)tr_4s + 1} \right] \left(\frac{k_p k_1}{1 - k_2 e^{-t_d s}} \right) \left[\frac{(tr_0s + 1)k_o}{tr_0s} + \frac{k_5 pH_L(s)}{tr_3s} \right] + t_d s} \quad (15)$$

$$\lim_{s \rightarrow 0} [s pH_i(s)] = 0 \quad \left| \quad pH_L(s) = \frac{1}{s} \right. \quad (16)$$

$$\lim_{s \rightarrow 0} [s pH_i(s)] = 0 \quad \left| \quad pH_L(s) = \frac{1}{s^2} \right. \quad (17)$$

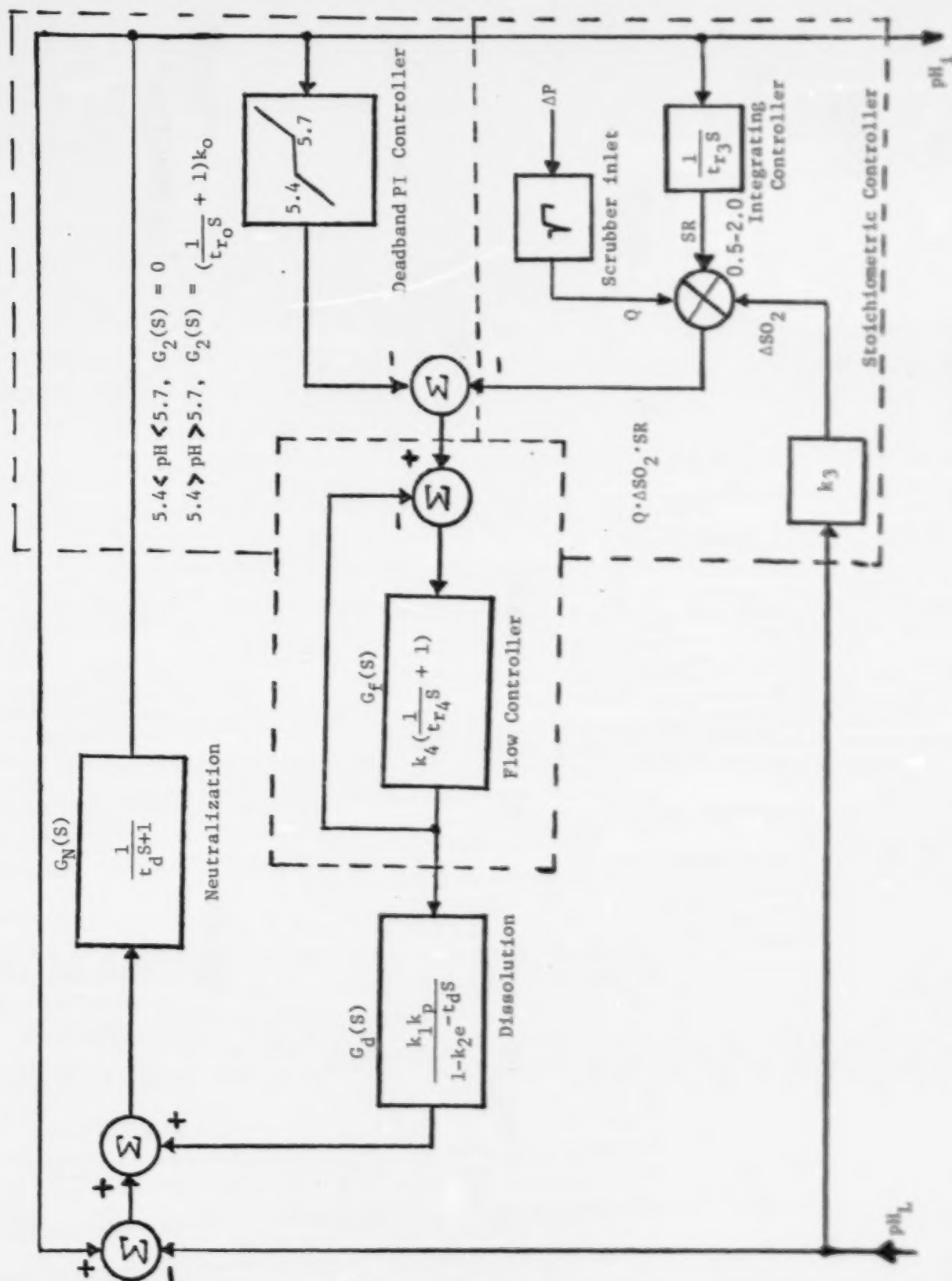


Figure 12. Combined Stoichiometric and pH Control Model

Parameter	Definition
$k_p = 0.1$	Slurry process gain
$k_o = 0.5$	Deadband controller gain
$k_1 = 0.037$	Limestone dissolution constant
$k_2 = 0.963$	Limestone recirculation constant
$k_3 = 2000 \frac{\text{ppm SO}_2}{\text{pH}}$	SO ₂ to pH conversion factor
$k_4 = 4$	Flow controller gain
$k_5 = k_3 \cdot Q$	SO ₂ mass flow multiplier
$t_d = 10 \text{ min}$	Hold tank residence time
$tr_o = 5 \text{ min}$	Deadband controller reset rate
$tr_3 = 30 \text{ min}$	Integrating controller reset rate
$tr_4 = 3 \text{ min}$	Flow controller reset rate

Table 8. Parameter Values For Model

SECTION 8

PH CONTROL EXPERIMENTAL RESULTS

Installation of the pH control system on the TVA Shawnee pilot limestone turbulent contact absorber (TCA) scrubber at Paducah, Kentucky, was completed in July, 1979, as shown by Figure 13. A narrow range 4.5 - 6.5 pH probe was utilized, an Electrofact Model 135G/Z, with mechanical probe wiping action and a pressurized reference electrode immersed in the scrubber slurry at the absorber inlet. A 4-pole 1.0 Hz sharp-cutoff Butterworth lowpass filter was inserted in the pH measurement signal circuit to prevent process noise from entering the control system to insure well-behaved control dynamics.

Start-up activities included online tuning of the batching and flow controllers obtained from observation of the subsidence of the pH and limestone flow controlled variables, following oscillation induced by excessive controller gain. With the integral term turned off the proportional band is adjusted until consecutive oscillation cycles exhibit an amplitude decay of 1-to- $\frac{1}{2}$, thereby defining the appropriate controller gain. The integral term is then defined equal to the oscillation time period for a minimum-error integral setting. Table 9 shows the controller tuning adjustments thereby obtained, which are in conformance with the classical methods of Ziegler and Nichols.⁴ Note that the flow controller gain is six times that of the batching controller, which satisfies the cascade

⁴Ziegler, J.G. and N.B.Nichols, Optimum Settings for Automatic Controllers, Trans. ASME, 64:759, 1942.

control requirement that the secondary loop have at least four times the gain of the primary loop.

Controller	Proportional Band	(Gain)	Reset Rate
Batching	200%	(0.5)	5 minutes
Flow	33%	(3)	3 minutes

Table 9. PH Control Tuning Adjustments

The scrubber neutralization reaction gain K_p was then experimentally determined at three slurry pH values; 5.3, 5.55, and 5.8. This component of the test plan was obtained by bumping the limestone feedrate in 20% steps 1.6 gpm (6 liters/minute) from the above steady-state pH values and noting the change ΔpH . The ratio of these values ($\Delta pH / \Delta \text{limestone feedrate}$) defines K_p , which is graphically presented versus slurry pH in Figure 14. The significance of this curve is imperative to successful and responsive limestone scrubber reagent control. The average value of reaction gain obtained over the practical range for pH control was $0.35 \Delta pH / \Delta \text{gpm}$ ($0.1 \Delta pH / \Delta \text{liters/minute}$). The closed-loop gain of this pH control system is determined by application of equation (14), where the primary controller has a pH span of 2.0 units (4.5 - 6.5). The closed-loop gain of the secondary flow loop is always 1.0, because it effectively relays the primary controller output signal and actuates the 30 liter/minute maximum dense limestone delivery system. The 0.7 gain obtained provides unconditional closed-loop stability.

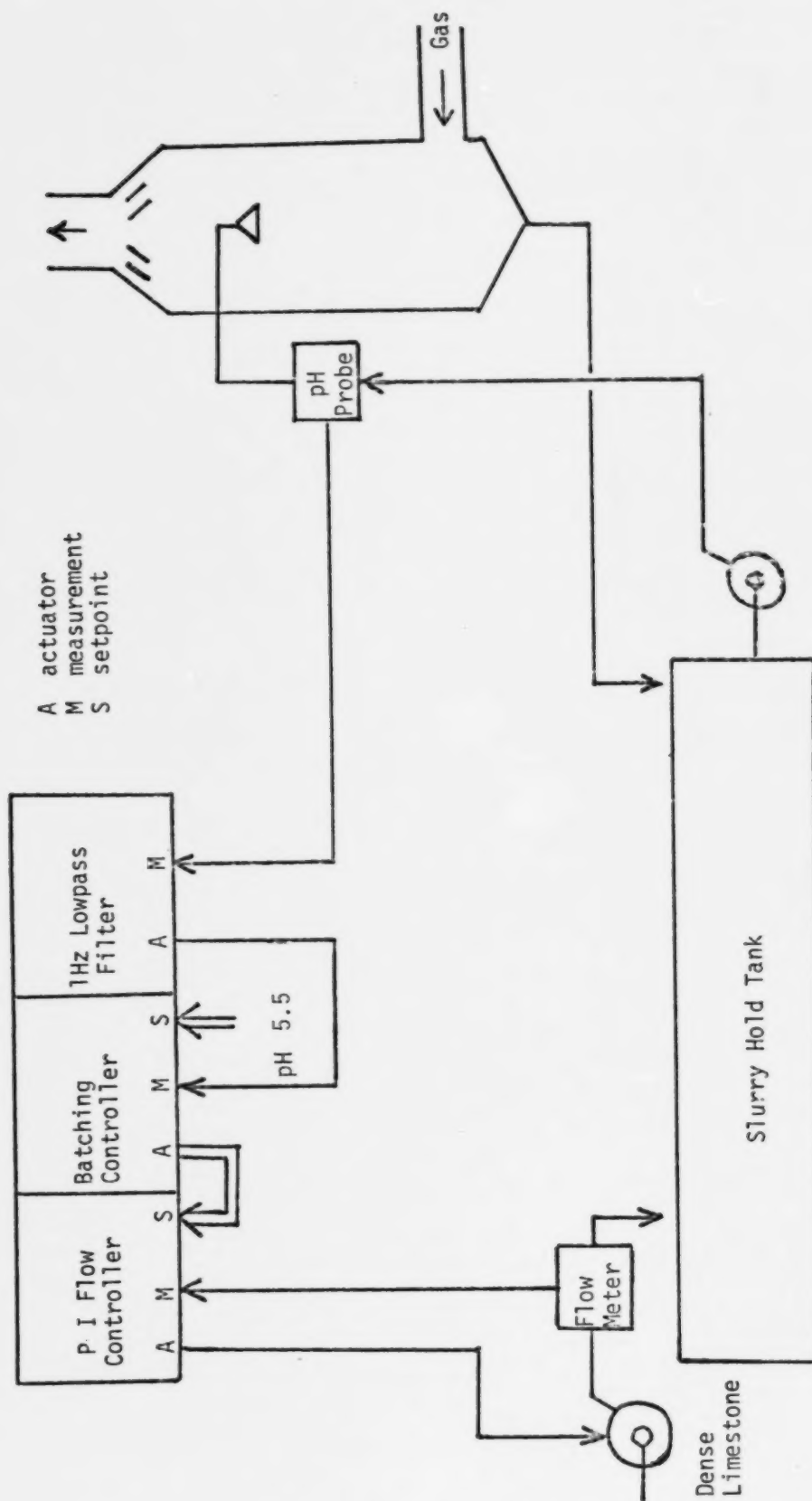


Figure 13. Slurry pH Control System Tested

$$\begin{aligned}
 \text{closed-loop gain} &= (\text{Primary gain})(\text{Flow gain})(\text{Reaction gain})(\text{Scaling}) \quad (14) \\
 &= (0.5) \cdot (1.0) \cdot (0.35 \frac{\Delta \text{ pH}}{\Delta \text{ gpm}}) \left(\frac{8 \text{ gpm max}}{2.0 \text{ pH span}} \right) \\
 &= 0.7
 \end{aligned}$$

Figure 15 describes the relationship of liquor pH to stoichiometric ratio (the inverse of limestone utilization) obtained by Bechtel at Shawnee which shows a remarkable correlation with the reaction gain of Figure 13. Parameters related to pH control of slurry limestone feedrate are summarized by Table 10, in which all three parameters exhibit limiting conditions above a slurry pH of 5.7 - 5.8. PH is also useful for predicting the utilization disorders of soft scaling at higher pH values (>5.8) and pH blinding at lower pH values (<4.5). Consequently, a limestone slurry pH setpoint in the vicinity of 5.6 maximizes SO₂ removal (70% range without organic acid buffering) and limestone utilization efficiency (typically 85%) with freedom from both internal scaling and blinding effects. This is also the upper range of slurry pH which offers a consistently high reaction gain for responsive and stable feedback pH control.

Parameter	Definition
Reaction Gain	$\Delta \text{ pH} / \Delta \text{ limestone feedrate}$
Stoichiometric ratio	moles Ca / moles SO ₂
Sulfur removal	SO _{2i} - SO _{2o}

Table 10. PH Related Slurry Parameters

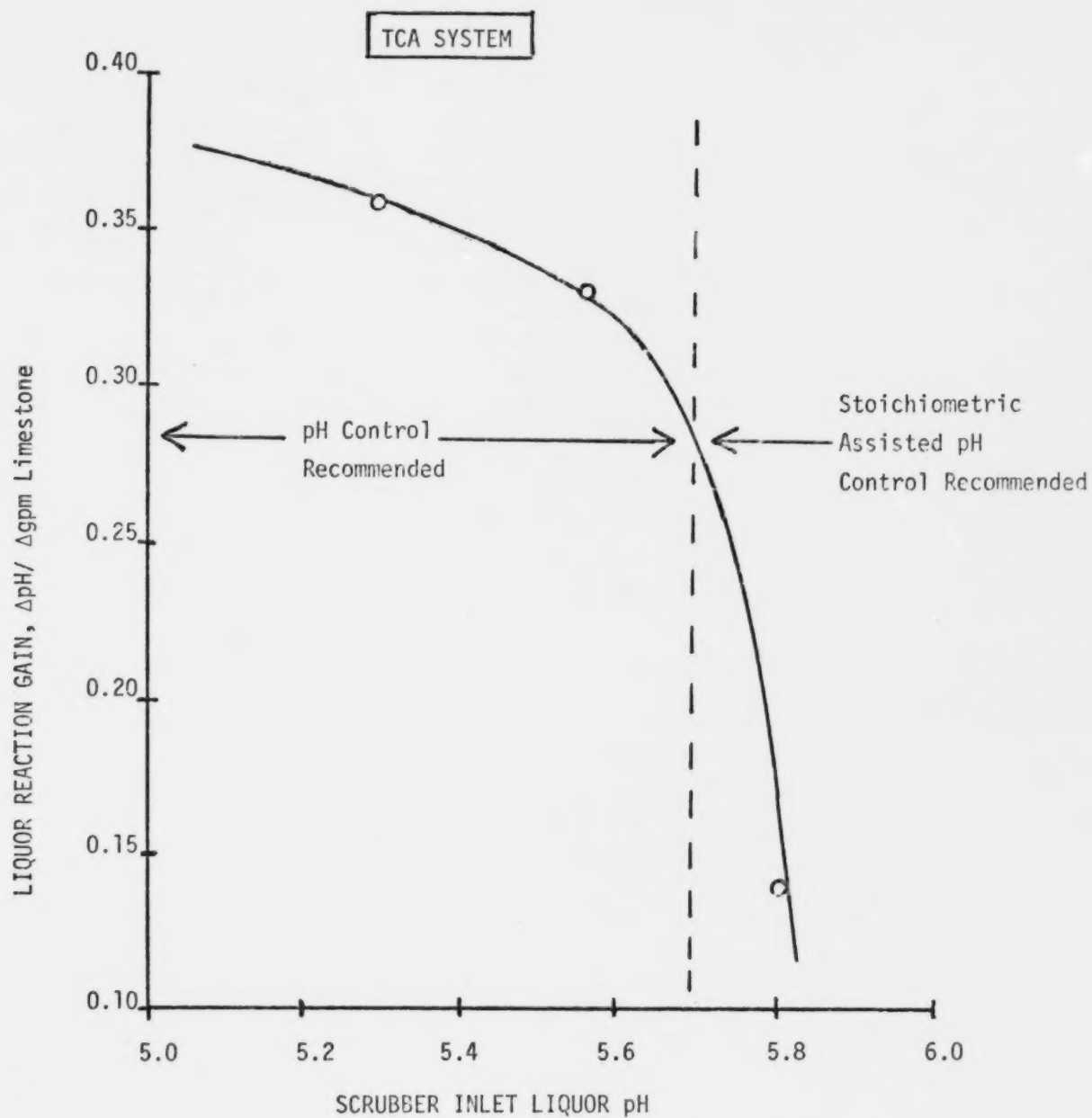


Figure 14. Reaction Gain Versus **Slurry pH**(after Garrett)

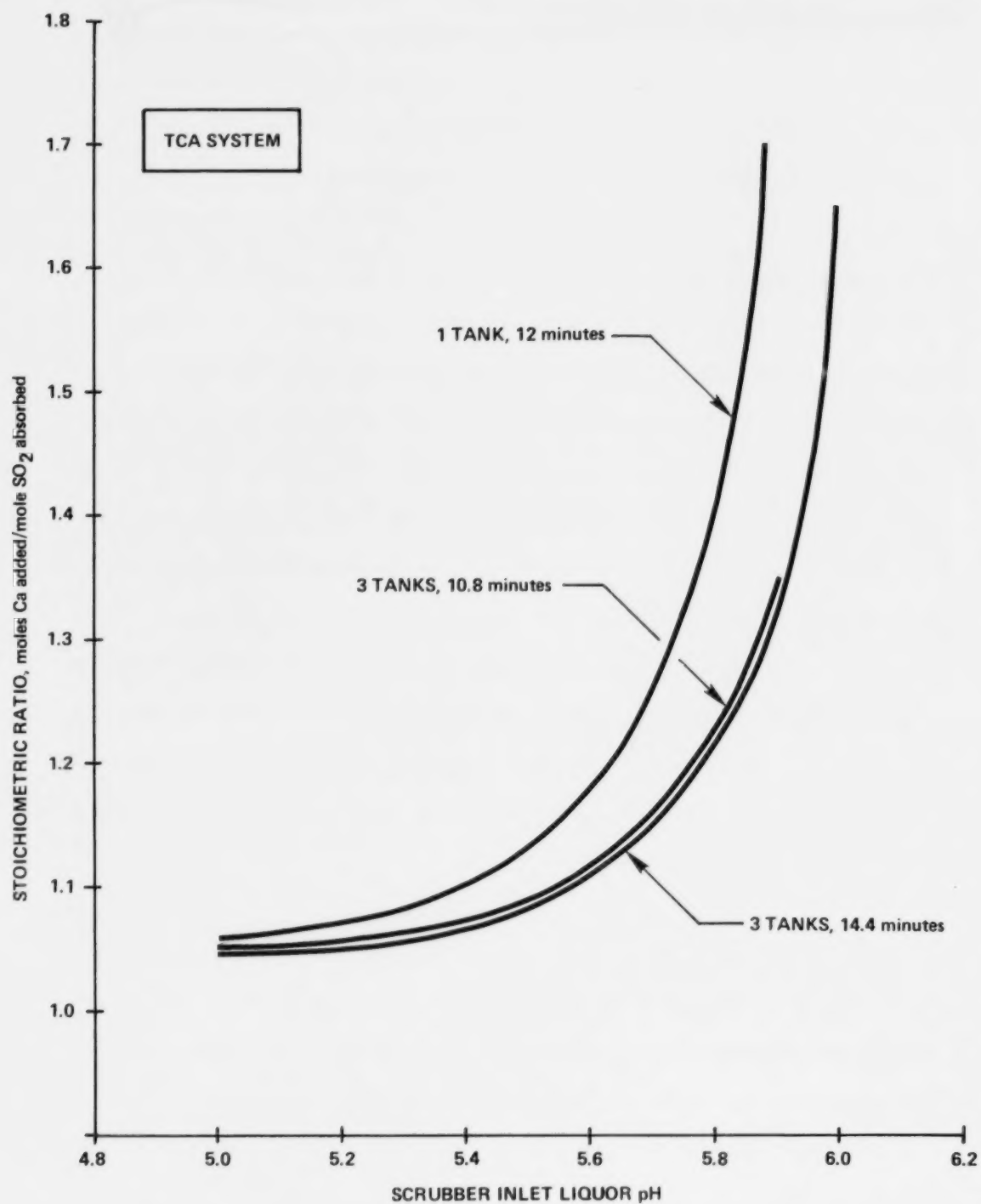


Figure 15. The Effect of Scrubber Inlet Liquor pH (after Bechtel)

Planned experiments were conducted from July 26th to the 30th, 1979. Figure 16 presents a recorded segment of the pH and limestone flow controlled variables obtained July 28th and 29th. It is notable that under stable scrubber conditions between 6 PM and 9 PM pH lined out at its setpoint within ± 0.01 pH unit, which verified the performance capability of this control system design for linear continuous control without erratic behavior such as on-off cycling. Successful pH control was also maintained during an unplanned limestone pipe restriction evident in the non-symmetrical flow between midday and 4 PM.

SECTION 9

STOICHIOMETRIC ASSISTED PH CONTROL EXPERIMENTAL RESULTS

Use of feedforward control elements to determine limestone feed-rate is generally construed as stoichiometric control. However, in the absence of measurement of liquid-phase reactions, fly ash alkalinity and chloride additions result in feedrate error because of violation of the physical chemistry assumptions described in Section 5. Slurry pH measurement represents an agglomerate of these liquid-phase reactions and therefore is advantageous in a reagent control system. Parallel operation of feedforward stoichiometric and pH feedback offers improved controllability because of utilization of both inlet gas and liquid-phase parameters, plus the reliability of independent limestone addition systems. Cascade flow control is included to maintain accurate control in the event of limestone addition system disturbances.

Figure 17 describes the stoichiometric assisted pH limestone feed-rate control system tested on the Shawnee TCA scrubber in December, 1980. This mechanization realizes all of the features included in the discussion of Section 7 and Figure 10, plus an integrating-only (memory) feedback controller to adjust the feedforward gain (SR) of the primary stoichiometric control loop (decreasing SR for increasing pH). This adjustment is made in response to offset between the slurry pH measurement and reference pH setpoint. A deadband PI controller provides independent backup control for pH excursions outside a preselected and setpoint-centered deadband, for example 5.3 - 5.7 pH, to accommodate major scrubber disturbances. Equations (18) and (19) present the

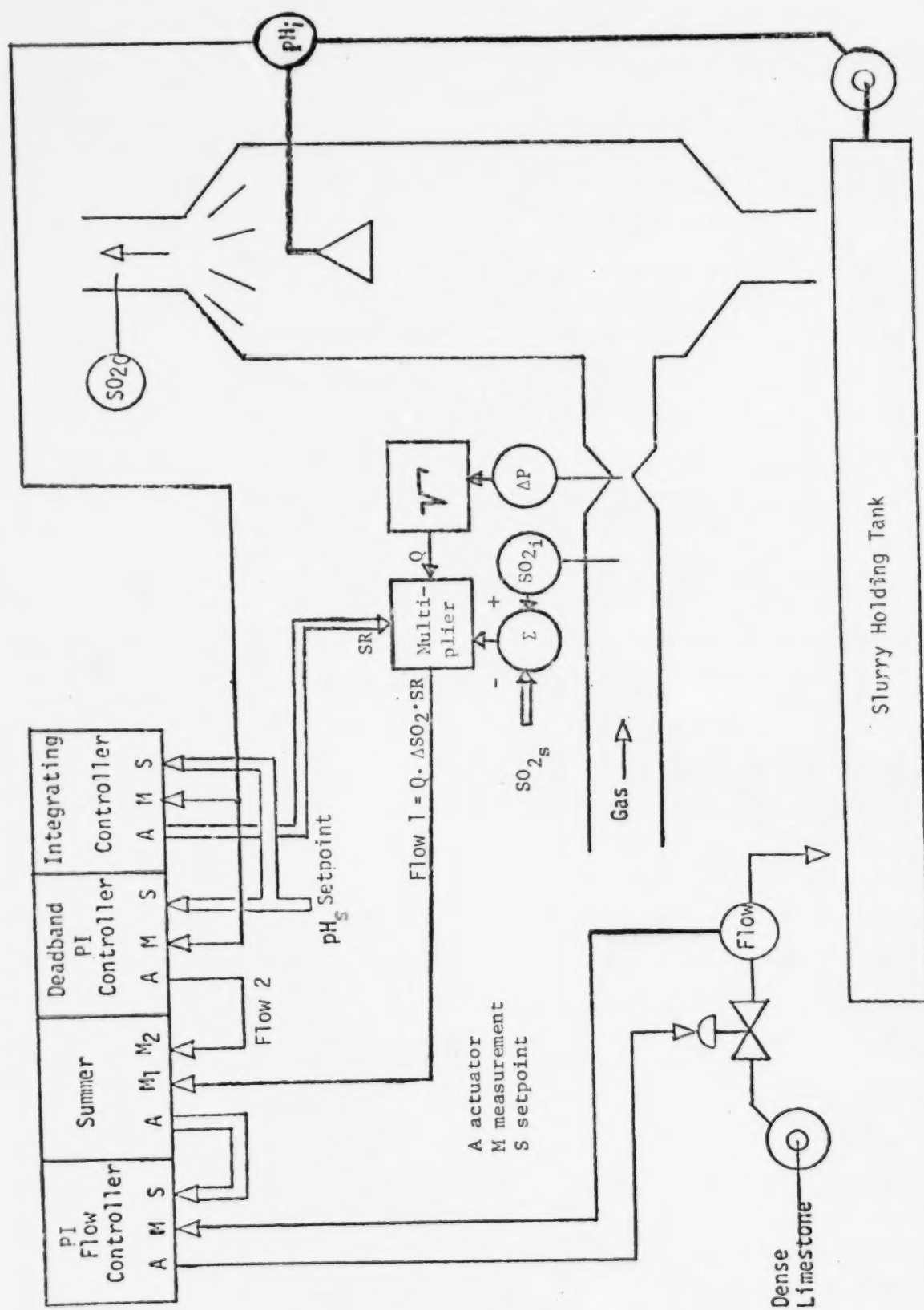


Figure 17. Steichiometric Assisted pH Control System Tested

relationships which represent these equally weighted parallel limestone addition signals.

$$\begin{aligned}
 \text{Flow 1} &= (\text{Inlet Variables})(\text{Feedforward gain})(\text{Scaling}) & (18) \\
 &= (\Delta\text{SO}_2 \times Q) (\text{SR}) \left(\frac{\text{moles Ca/ moles SO}_2}{\text{limestone purity} \times \text{slurry lb/gal} \times \text{density} \times \text{specific gravity}} \right) \\
 &= (\Delta\text{SO}_2 \times Q) (\text{SR}) \left(\frac{100/64}{0.97 \times 8.3 \times 0.6 \times 1.61} \right) \\
 &= \Delta\text{SO}_2 \times Q \times \text{SR} \times 0.2 \text{ (gpm)}
 \end{aligned}$$

$$\begin{aligned}
 \text{Flow 2} &= (\text{pH}_S - \text{pH}_i) (\text{Controller gain}) (\text{Scaling}) & (19) \\
 &= (\text{pH}_S - \text{pH}_i) (0.5) \left(\frac{8 \text{ gpm max}}{2.0 \text{ pH span}} \right) \\
 &= (\text{pH}_S - \text{pH}_i) \times 2.0 \text{ (gpm)}
 \end{aligned}$$

Installation of this control system on the TCA scrubber was accomplished in November, 1980, and preliminary testing begun the first week of December. This was interrupted due to boiler outage of one week duration, and then proceeded in competition with other scheduled tests. Consequently, in lieu of extended factorial tests experimental proofing for the simulation of feedforward assisted pH feedback control reported as Run (a) in Figure 5 was undertaken to validate this control method within the time available. In accordance with the simulation exercise a setpoint of 5.5 pH was elected and 1000 ppm (0.2 pH) load changes approximated by simultaneous 10,000 cfm flue gas and 600 gpm slurry changes. A deadband controller setpoint of 5.5 ± 0.2 pH properly provided zero Flow 2 under these conditions,

with feedforward control determining limestone addition through Flow 1. The controller tuning adjustments used are shown in Table 11.

Controller	Proportional Band	(Gain)	Reset Rate
SR	-	(1.2 initial)	-
Deadband	200%	(0.5)	5 minutes
Flow	33%	(3.0)	3 minutes
Integrating	—	—	15 minutes

Table 11. Controller Tuning Adjustments

A recording of SR and slurry pH are shown in Figure 18 with tests beginning December 23rd at a ΔSO_2 of 2000 ppm, gas rate of 20,000 cfm, and slurry flow rate of 600 gpm. The initial test phase demonstrates the ability of the control system to maintain a pH setpoint of 5.5 within 0.04 pH unit under steady-state scrubber operating conditions. A feedforward gain equivalent to an SR of 1.9 was required to achieve this. At 2:10 PM a load decrease was introduced by increasing the gas rate to 30,000 cfm. This major scrubber disturbance resulted in limit cycling of SR to its maximum 2.4 value, and a negative slurry pH half-cycle of 50 minutes duration and 5.35 pH minima. The following 90 minute positive pH half-cycle was accompanied by zero SR and a 5.61 pH maxima. Excess entrained limestone was depleted during this interval for the slurry to return to its 5.5 pH setpoint. At 4:10 PM scrubber load was decreased to the initial conditions of 20,000 cfm and 600 gpm,

and SR remained zero as a consequence of excess entrained limestone for these conditions. One extended 120 minute positive pH half-cycle was obtained with a maxima of 5.66 pH in returning to the 5.5 pH setpoint. Throughout this test the deadband controller output was zero since all slurry pH excursions were within its 5.5 ± 0.2 pH deadband. Table 12 provides a comparison of these experimental results with the simulation exercises of Figure 5.

A significant conclusion provided by this effort is that a potential choice of methods exist for limestone feedrate control, with selection based upon that portion of the reaction gain curve of Figure 14 over which operation is intended. For high reaction gains, available up to slurry pH values of 5.7 - 5.8, a batching cascade pH control system is recommended for continuous linear and stable control of limestone addition. This mechanization may also be realized with the stoichiometric assisted pH control system by setting SR to zero disabling Flow 1, and removing the deadband (5.5 ± 0.0 pH) from the deadband controller enabling Flow 2 full time. The additional complexity of stoichiometric assisted pH control of limestone feedrate is primarily warranted for scrubber operation at low reaction gains, experienced at slurry pH values above 5.8. Lower slurry reaction gain is compensated by the additional responsiveness inherent in the feed-forward control elements of this mechanization.

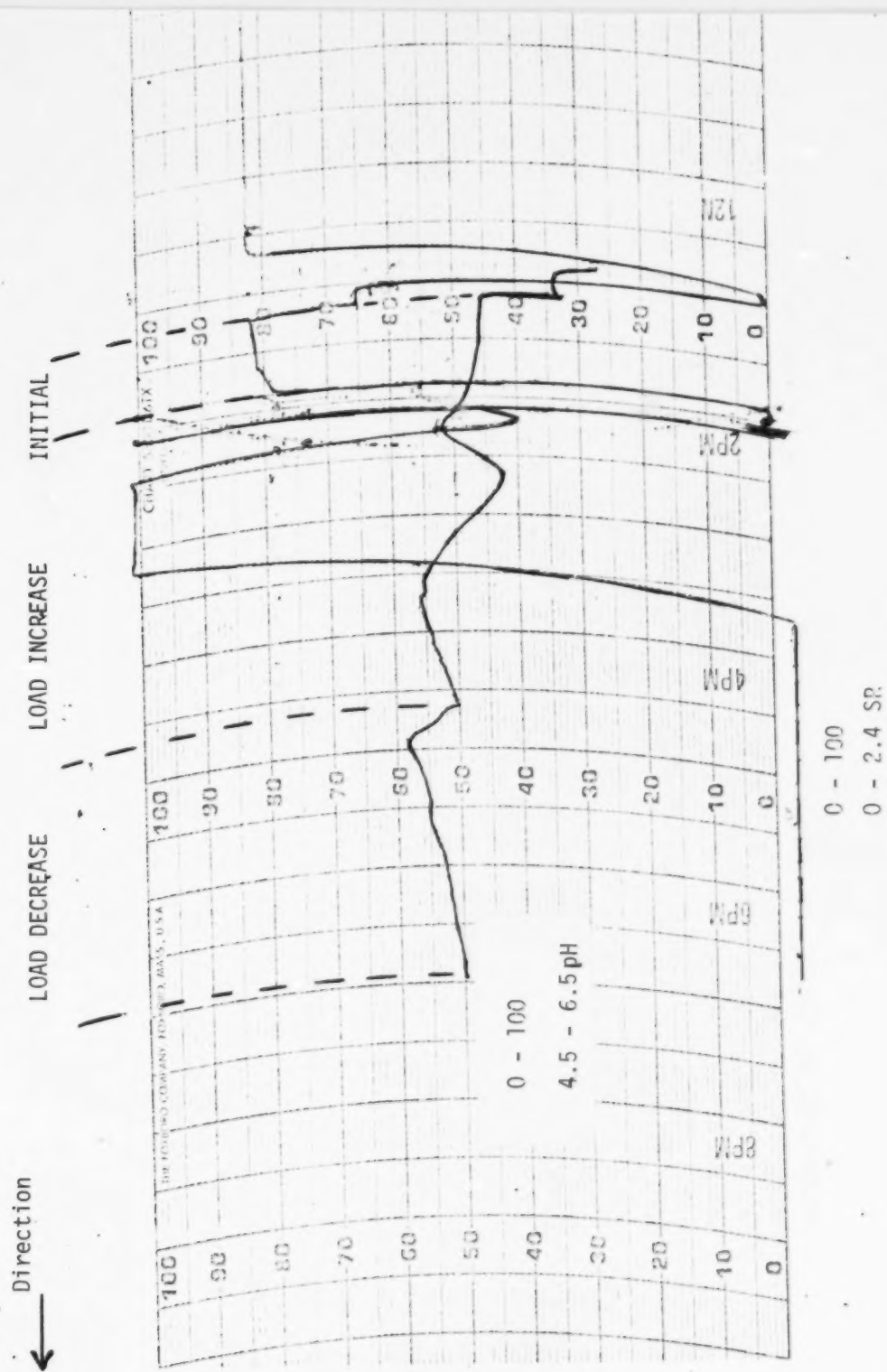


Figure 18. Liquid pH and Adaptive SR

LOAD CHANGE	SIMULATION		TEST	
	Peak	Half-Cycle	Peak	Half-Cycle
1000ppm SO ₂ increase	5.45pH	60 min	5.35pH	50min
	5.56pH	100 min	5.61pH	90min
1000ppm SO ₂ decrease	5.63pH	180 min	5.66pH	120min

TABLE 12. Simulation Versus Test Comparisons

SECTION 10

SLURRY DENSITY CONTROL MODELING AND SIMULATION

It is essential to control the density of the recirculated limestone slurry in order to maintain an adequate number of crystallization sites for calcium sulfate precipitation, while simultaneously insuring that circulating sulfite species are insufficient to produce internal scrubber scaling. These requirements can be met by the regulation of the slurry solids to within a narrow variance, typically 15 percent by weight, aided by a slurry density measurement. The relationship between density and percent solids for a limestone slurry is defined by equation (20), where the latter is a more responsive controlled variable for automatic control purposes. In order to evaluate the dynamics of this control loop and arrive at a control design capable of good performance under anticipated scrubber operating conditions, a mathematical model of the slurry density control loop was developed and simulated analogous to the study for pH control of limestone feedrate. For consistency both studies were conducted with a slurry holding tank residence time of 10 minutes, and SO_2 load changes corresponding to 1000 ppm about 2500 ppm.

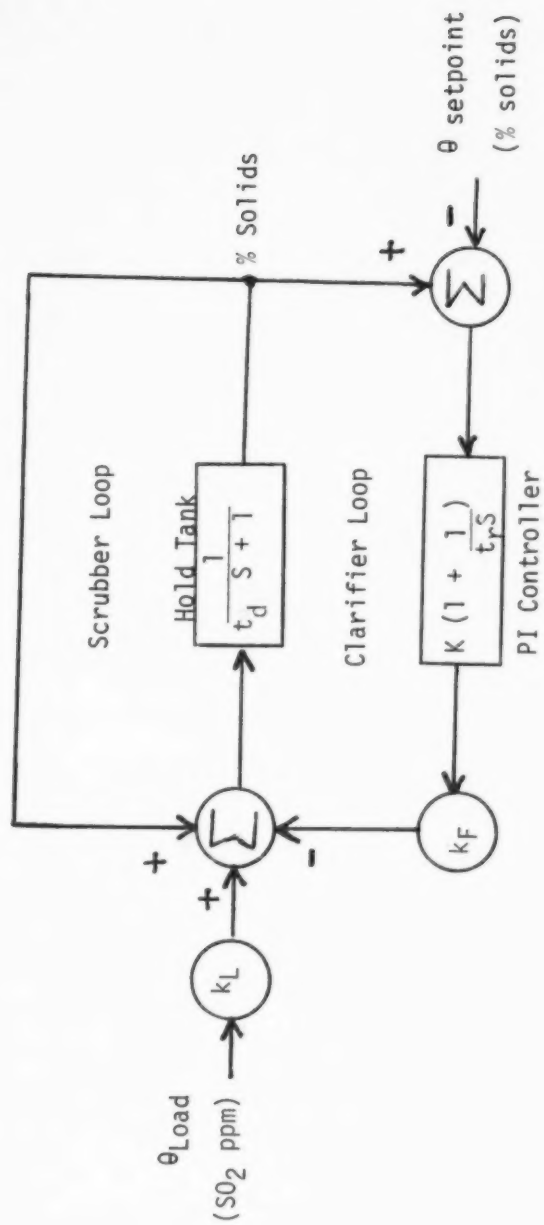
$$\text{percent solids} = 165 \left[1 - \frac{1}{\text{specific gravity}} \right] \quad (20)$$

The density control loop is modeled as the single-capacity process shown in Figure 19, whose time constant t_d is the 10 minute holding tank residence time and represents the crystallization time to a first approximation. The crystallization process gain k_L and clarifier loop gain k_F

shown in this figure were derived from data available for the TCA limestone scrubber at the TVA Shawnee facility as presented in equations (21) and (22). A verification of these constants was obtained by noting that k_L and k_F are in material balance for observed scrubber steady-state conditions of 2500 ppm SO_2 loading and a 95 liters per minute clarifier slurry sidestream at 15 percent solids by weight. Simulation runs of the slurry density control system are shown in Figure 20 under the control of both PI and P (integral term turned off) feedback controllers. Run (a) is for P control with a proportional band of 10 percent which produces a 7 percent steady-state error (1.05% solids) in the controlled variable following a load change equivalent to 1000 ppm. Consequently, a P controller is not capable of automatically returning the slurry percent solids to its setpoint following a load change. Further investigation determined that a PI controller with a proportional band of 5 percent and a 10 minute reset rate reduces steady-state error to 1 percent (0.15% solids) following load changes as shown in Run (c). This is the recommended density controller and is shown in the slurry density control system of Figure 21. Documentation of this simulation exercise follows.

$$\begin{aligned}
 k_L &= \frac{\text{Limestone Addition} \cdot \text{Fly ash}}{\text{Tank Volume} \cdot \text{Nominal } \text{SO}_2 \text{ Loading}} & (21) \\
 &= \frac{13.6 \text{ liters} \cdot 1.5}{45,400 \text{ liters} \cdot 2500 \text{ ppm}} \\
 &= 0.011\%/1000 \text{ ppm-minute}
 \end{aligned}$$

$$\begin{aligned}
 k_F &= \frac{\text{Clarifier Flow} \cdot \% \text{Solids} \cdot \text{Efficiency}}{\text{Tank Volume}} & (22) \\
 &= \frac{190 \text{ liters/minute} \cdot 15\% \cdot 0.9}{45,400 \text{ liters}} \\
 &= 0.054\%/minute
 \end{aligned}$$



$$k_L = 0.011 \text{ \%solids/1000 ppm-min}$$

$$k_F = 0.054 \text{ \%solids/min}$$

Figure 19. Slurry Density Control Model.

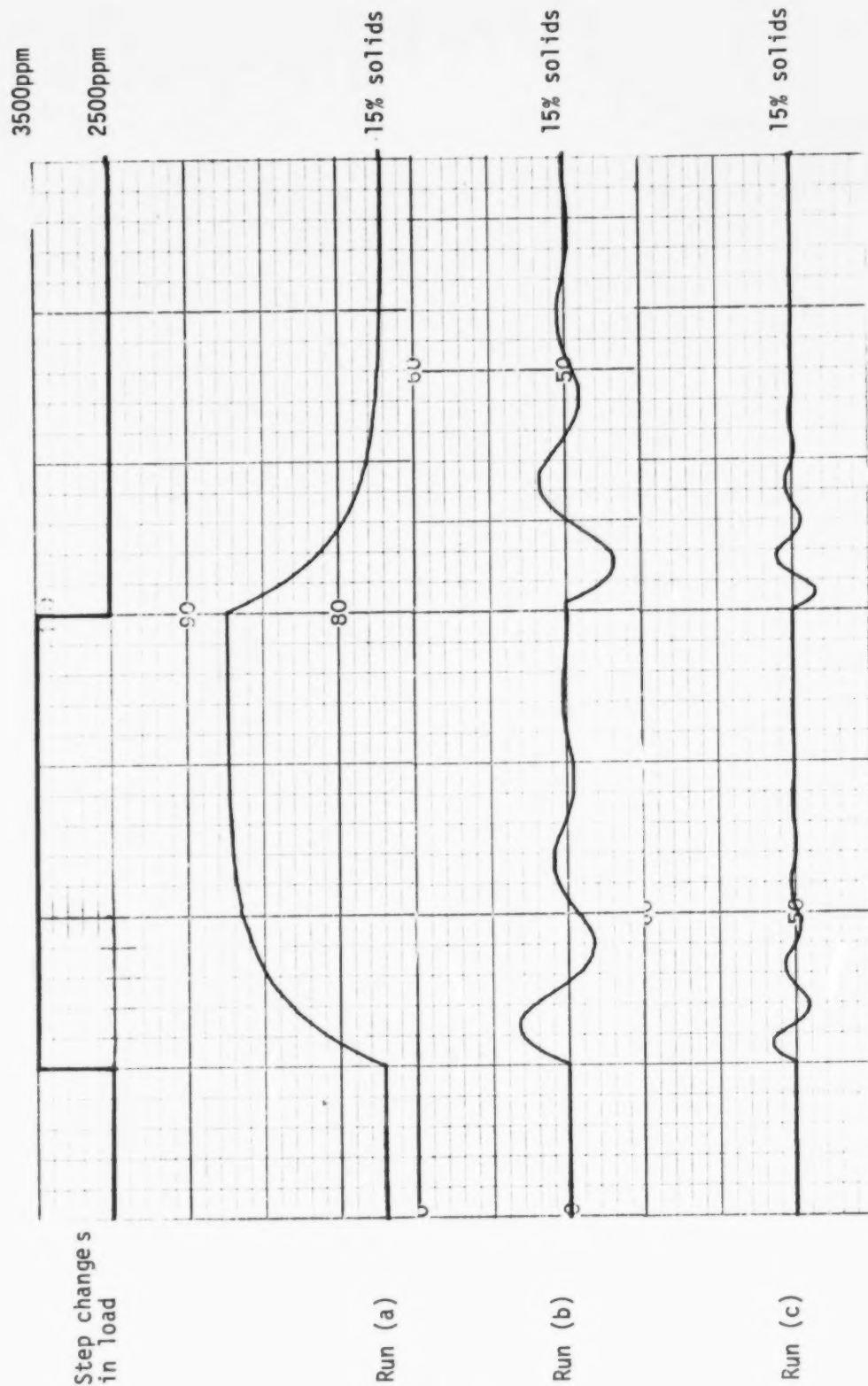


Figure 20. Density Control Simulation Runs

Legend: 100 minutes/horizontal division
0.1% solids/vertical division

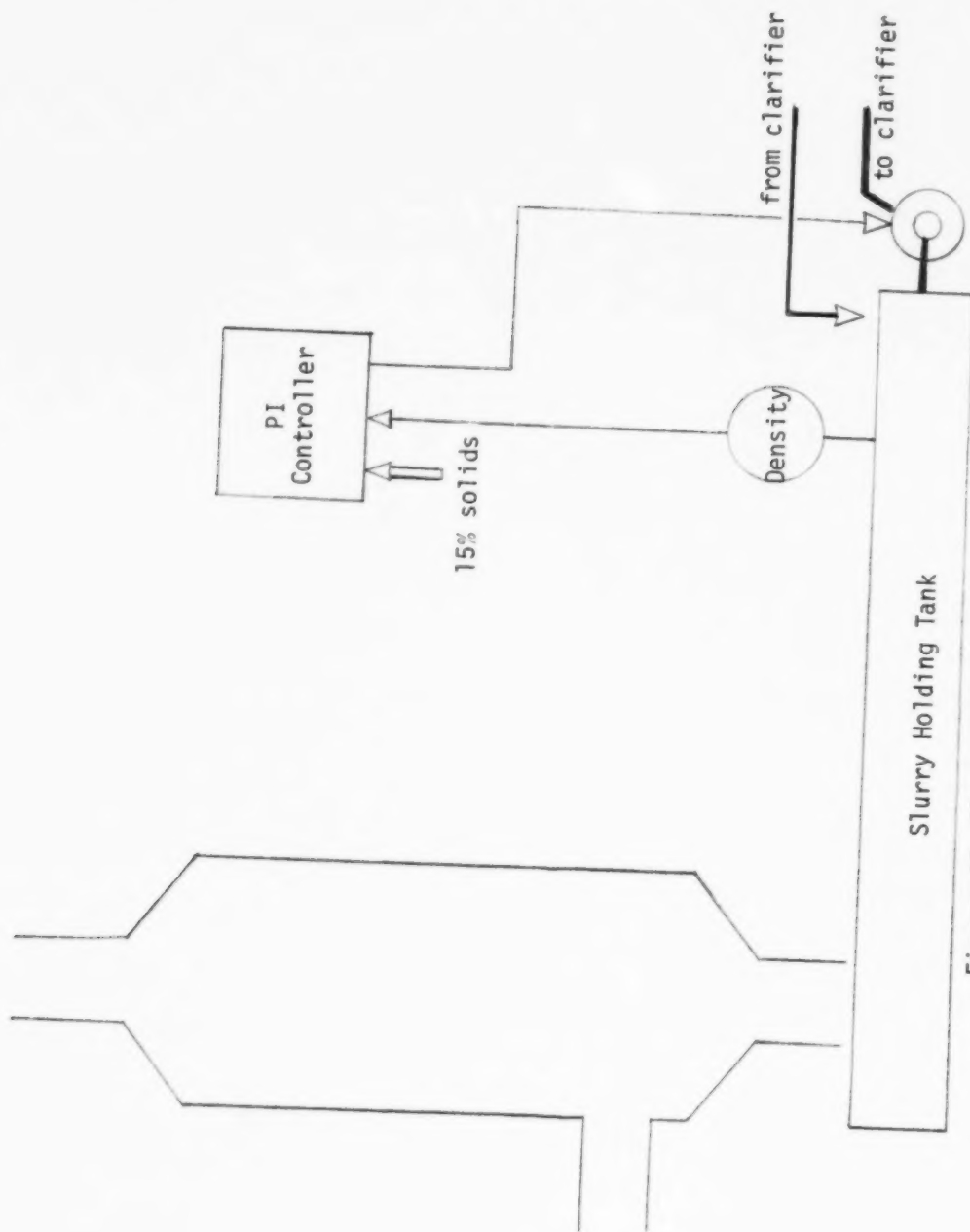


Figure 21. Slurry Density Control System

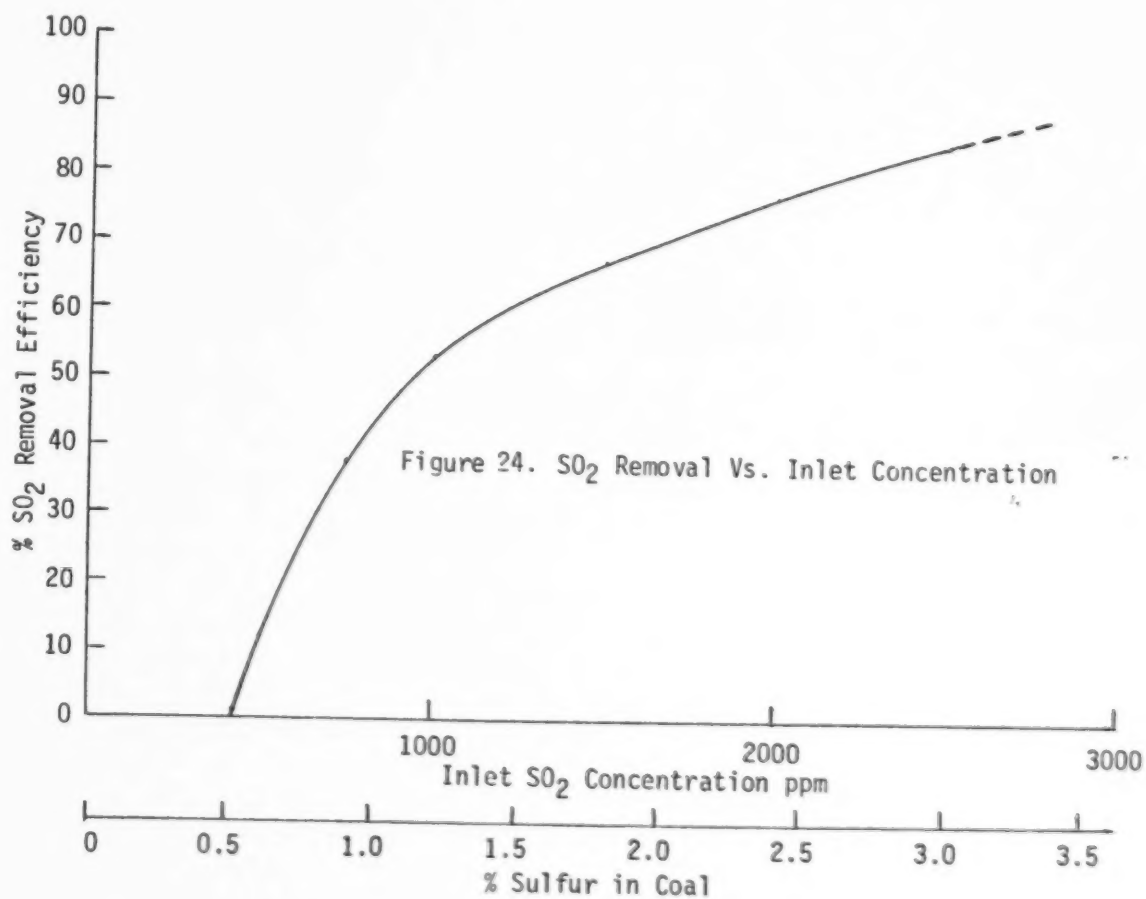
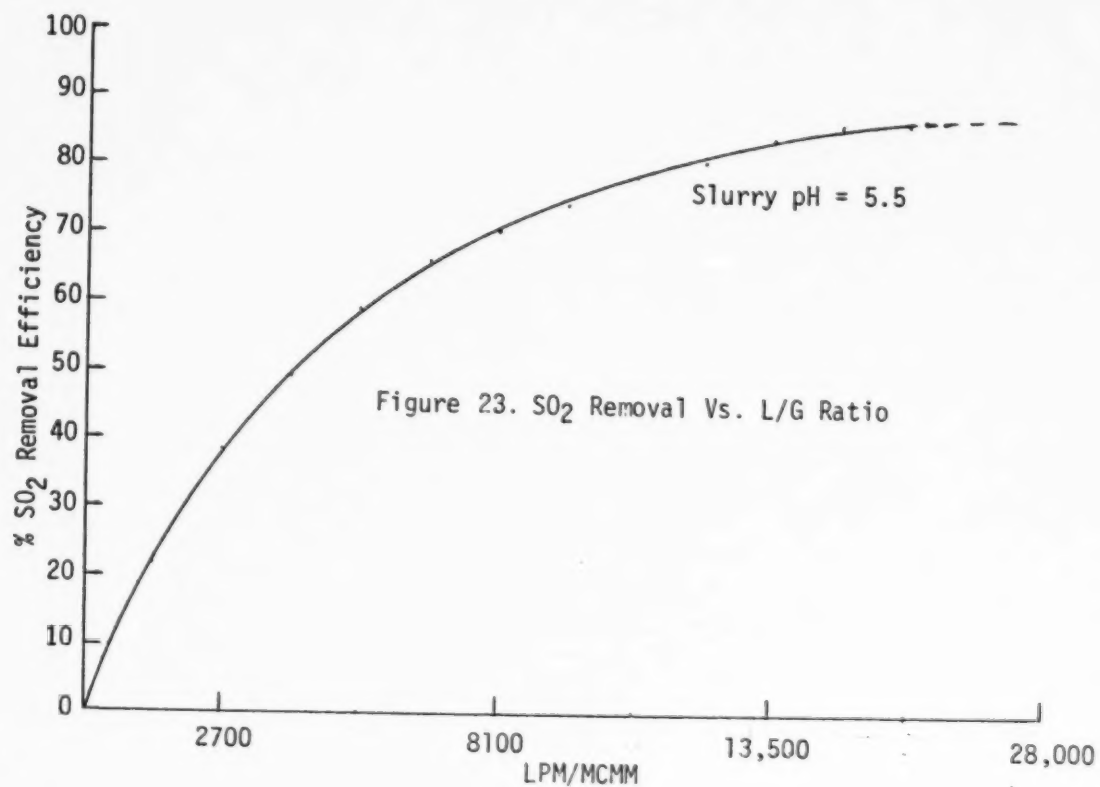
To insure consistency with the simulation performed for the pH-controlled limestone addition system, 1 second of computer run time again corresponds to 10 minutes of the actual process and the same 2500 ± 1000 ppm SO_2 load changes applied. The simulated controller output is limited between zero and 100 percent by the limiter circuits associated with amplifiers 7 and 14 of Figure 22. Potentiometer 13 permits adjustment of the limiting point, and potentiometer 10 the slurry solids 15 percent baseline value. Variable X_s represents the slurry solids controller setpoint, X_E the error signal, and X_p the output actuation signal. The integral time t_r is introduced by potentiometer 3 and integrator 3, and proportional gain k_C by potentiometer 2 and amplifier 4. Potentiometers 6, 7, and 8 provide the SO_2 load changes represented by X_L . The scrubber loop is simulated by integrator 9 whose output X_O recirculates through multiplier X and potentiometers 12 and 14, with the appropriate load gain k_L product of 0.011. This is countered by the clarifier flow rate represented by $-X_F$.

SECTION 11

LIQUID/GAS CONTROL MODELING AND SIMULATION

It is acknowledged that the energy efficiency of scrubber operations can be improved without impairing performance by adjusting the absorber L/G ratio to reflect SO_2 loading and, hence, removal efficiency requirements. Energy savings result from reduced induced-draft fan horsepower and pumping requirements when the L/G ratio is not maintained at a higher than required value. For example, coordination of Figures 23 and 25 provide that an inlet SO_2 concentration of 2500 ppm requires 90 percent removal to maintain an exit SO_2 target of 450 ppm, and L/G ratio of 16,000 liters per thousand cubic meters of flue gas (organic acid slurry buffering not considered). The absorber pressure drop from this L/G per 1000 kg SO_2 is found to require 355 fan horsepower to overcome. If inlet SO_2 drops to 1000 ppm, however, requirements to maintain the same 450 ppm exit target decrease to 55% removal efficiency at 4000 liters/MCM (GPM/MCFM = 135LPM/MCMM) and one-half the fan power, or 177 horsepower. It is clear that this magnitude of energy savings (power x time), plus that provided from furloughing slurry recycle pumps, cannot be ignored in the management of the scrubber in comparison with operation at a continuous upper design L/G value of the absorber. Consequently, a model of the scrubber absorber process was derived and simulated to investigate its dynamics and to identify an optimum control scheme.

A mathematical model was developed for the L/G process and anticipated control configuration as described by Figure 26. SO_2



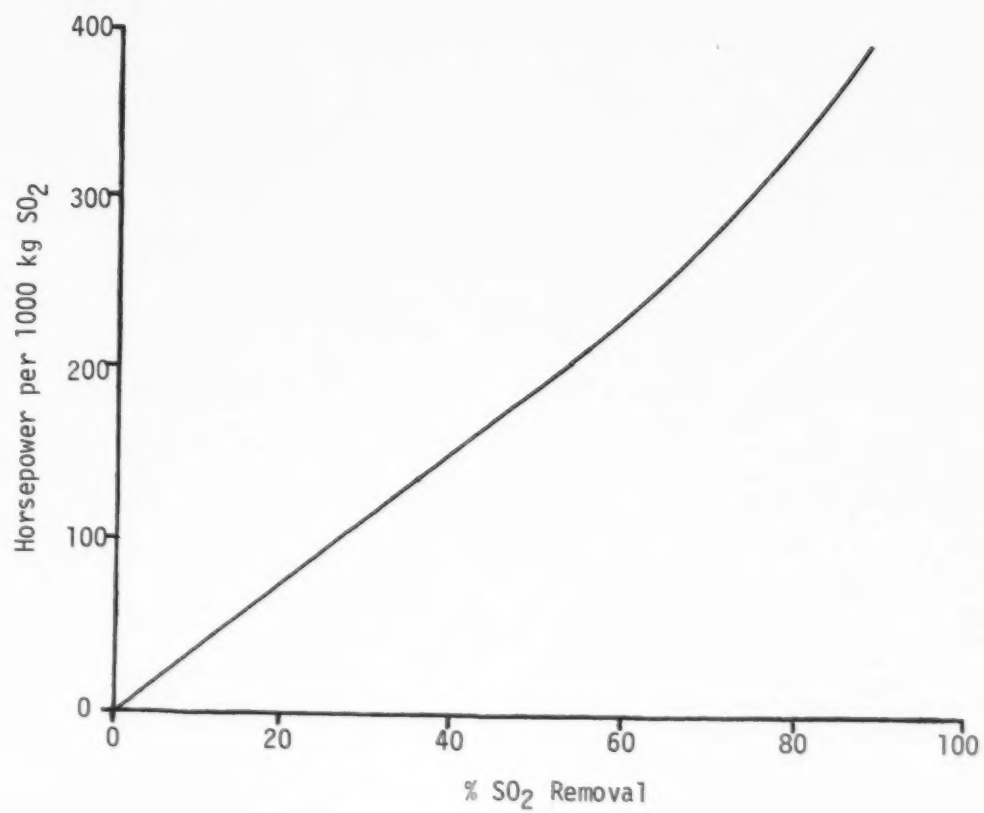


Figure 25. SO_2 Removal Vs. Fan Horsepower

absorption by a limestone slurry is modeled as a first-order reaction with a 0.5 minute time constant t_a and absorber gain $k_g \cdot \frac{\Delta \text{ppm}}{\Delta L/G}$. The L/G ratio is under the control of a feedforward law shown by Figure 27, and a pump-matrix selector based on Duhamel's integral controlling four recycle pumps of equal slurry delivery capability. The feedforward law is a nonlinear isopleth relating the variables of Figures 23 and 24 at a slurry pH of 5.5, and provides a contour of removal efficiency specifying the required L/G ratio for inlet SO_2 concentrations ranging from 450-3000 ppm. The absorber gain k_g is an average value obtained from Figure 20, and Table 13 provides a summary of the L/G control system model parameters.

An extension of this control system is a PI feedback controller measuring the exit SO_2 concentration and trimming the 450 ppm exit target feedforward setpoint. The rationale for this addition is to provide automatic compensation for variability in the SO_2 absorption process, reflected as error in the feedforward law, by adjusting the L/G ratio as required to maintain compliance with the exit target value. However, simulation disclosed that a feedback controller using measurements of scrubber outlet SO_2 creates unstable control system oscillations and therefore is not recommended. This instability is shown in Run (b) of Figure 28, and is caused by the controller hunting for an unavailable steady-state operating point. This occurs when the outlet SO_2 value exceeds the exit target setpoint and calls for an additional pump, which then depresses the outlet SO_2 value and calls for the furloughing of the same pump.

The solution is to utilize feedforward control without the feedback controller, and to accept whatever errors may accrue due to variations in the process, with the feedforward law nevertheless providing unconditionally stable L/G control and significant scrubber energy savings. Run (c) shows well-behaved system performance under feedforward-only control. The pump-matrix selector points are at 660 ppm inlet SO_2 intervals above the initial 450 ppm level, and are determined by a practical straight-line constant fit (FFL) to the feedforward law of Figure 27 as shown in Table 13. For increasing and decreasing inlet SO_2 levels up to 3000 ppm, the outlet SO_2 levels are reasonably well maintained within the 450 ppm exit target value, with an occasional excursion up to 600 ppm due to the coarseness of the available pump L/G steps. Conversely, intervals of excessive L/G ratio result in saturation effects and a minimum outlet SO_2 concentration in the vicinity of 100 ppm.

The recommended implementation of this feedforward L/G control system is described by Figure 29. The pump-matrix selector can be mechanized by a multiplying analog-to-digital converter which forms the product of the feedforward L/G command and flue gas volumetric flow Q , where Q is available from the material balance computer of Section 5, to generate the appropriate pump-selection quantization intervals. The feedforward law may be implemented as an arithmetic function described by its entry in Table 13 by simple linear circuitry. These results are for a full-scale scrubber with a slurry pH of 5.5, and demonstrate that SO_2 removal can be automatically maintained under varying scrubber load conditions with unconditionally

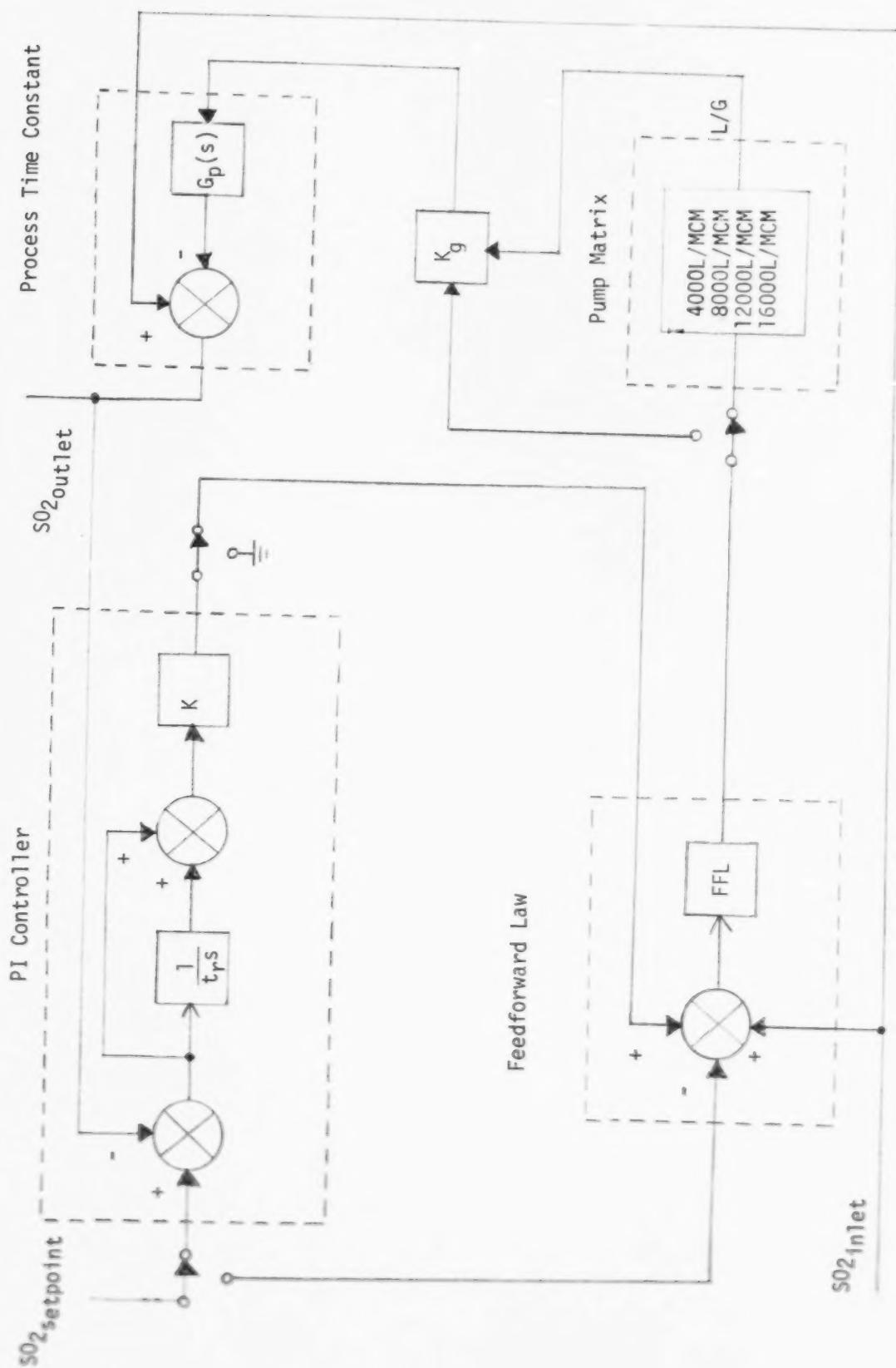


Figure 26. Feedforward L/G Control Model with Feedback

Element	Function	Description
Feedforward law	FFL	$6.75 \frac{\text{LPM/MCMM}}{\text{SO}_2 \text{ ppm}}$
Pump matrix	Duhamel's	$\sum_{i=1}^4 u(1 - \Delta \frac{\text{FFL} \cdot \text{SO}_2}{4000 \frac{\text{LPM}}{\text{MCMM}}}) \cdot 4000 \frac{\text{LPM}}{\text{MCMM}}$
Absorber gain	k_g	$0.16 \frac{\Delta \text{SO}_2}{\Delta \text{LPM/MCMM}}$
Absorber reaction $G_p(s)$	$\frac{1}{t_a s + 1}$	$\frac{1}{0.5s + 1}$
SO ₂ inlet	Ramp	3000 ppm
SO ₂ exit target	Constant	0 ppm
		450 ppm
Controller band	$\frac{100\%}{k}$	300%
Controller reset rate	$\frac{1}{t_r}$	0.5 per minute

Table 13. Summary Of L/G Control System Model

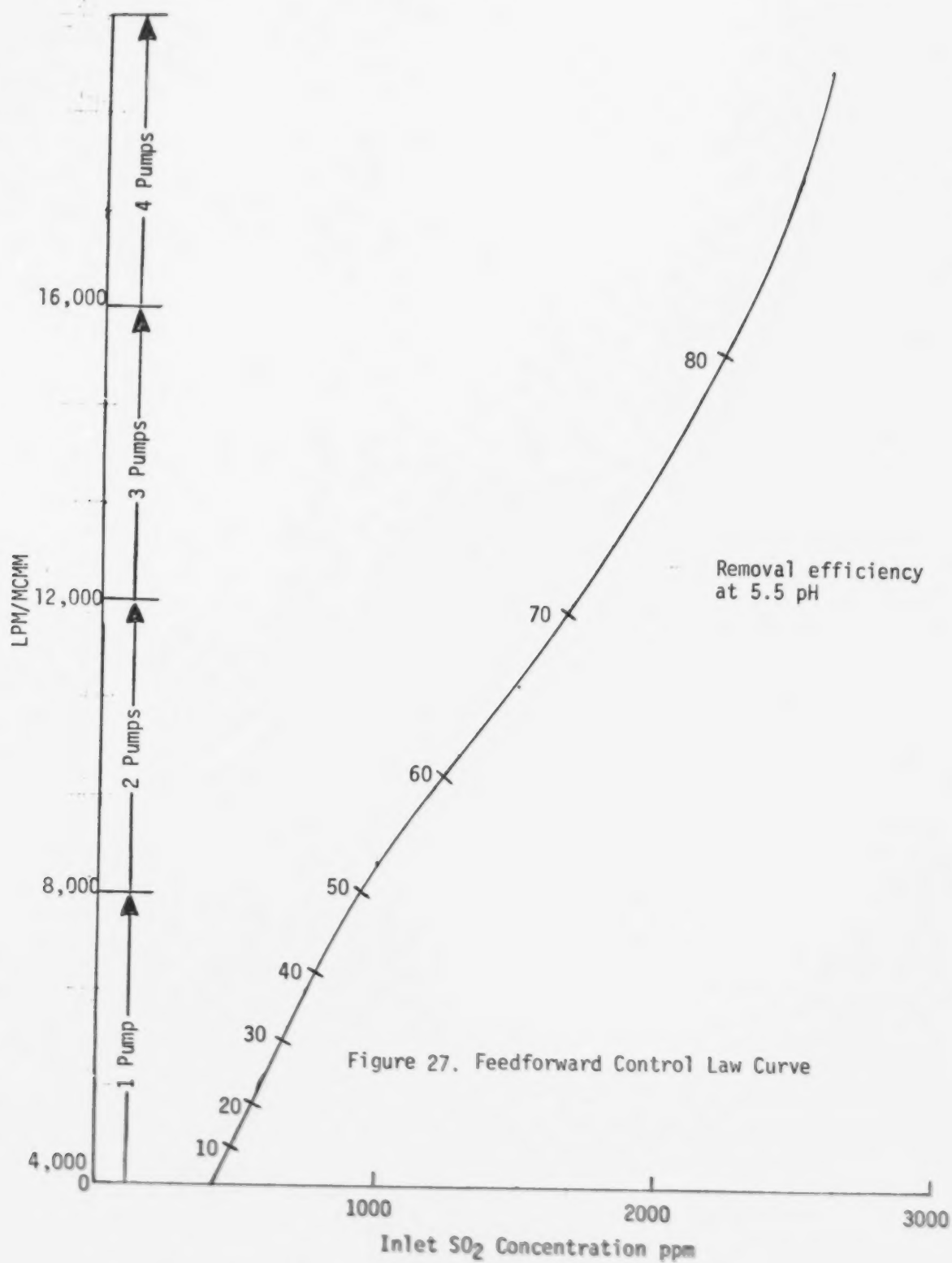


Figure 27. Feedforward Control Law Curve

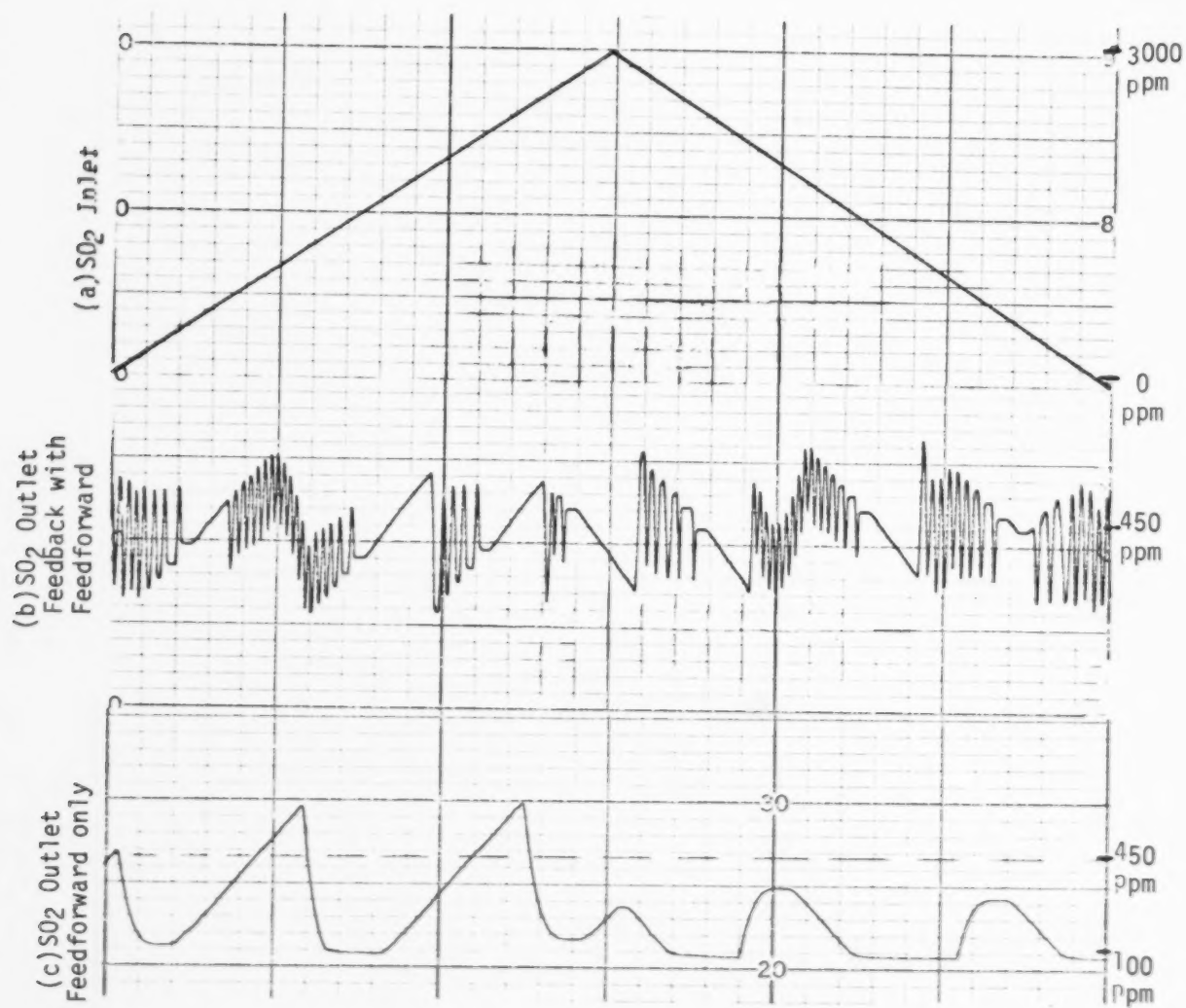


Figure 28. L/G Control Simulation Runs

Legend: 1 minute/horizontal division
 a. 150 ppm/vertical division
 b. 60 ppm/vertical division
 c. 60 ppm/vertical division

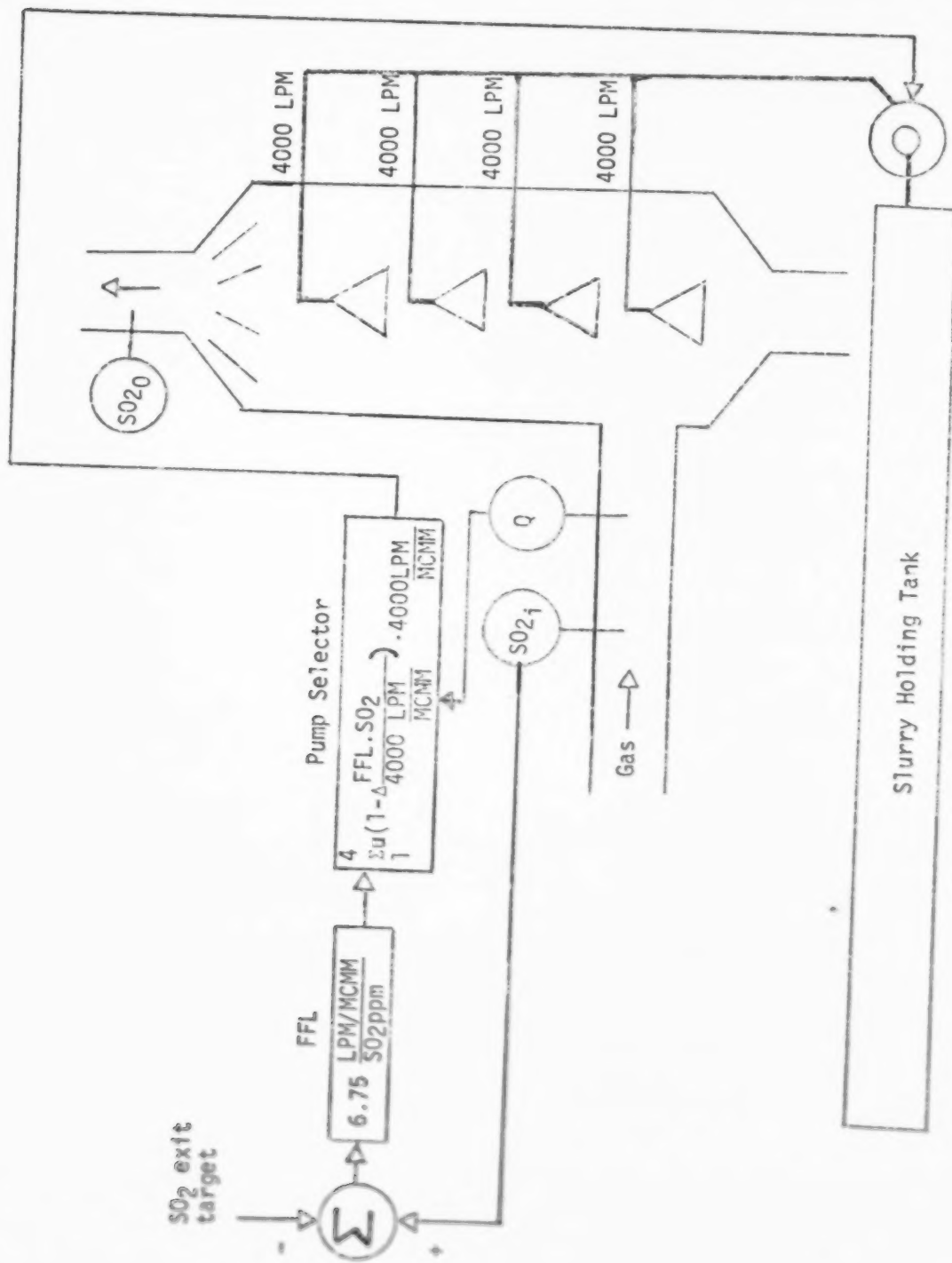
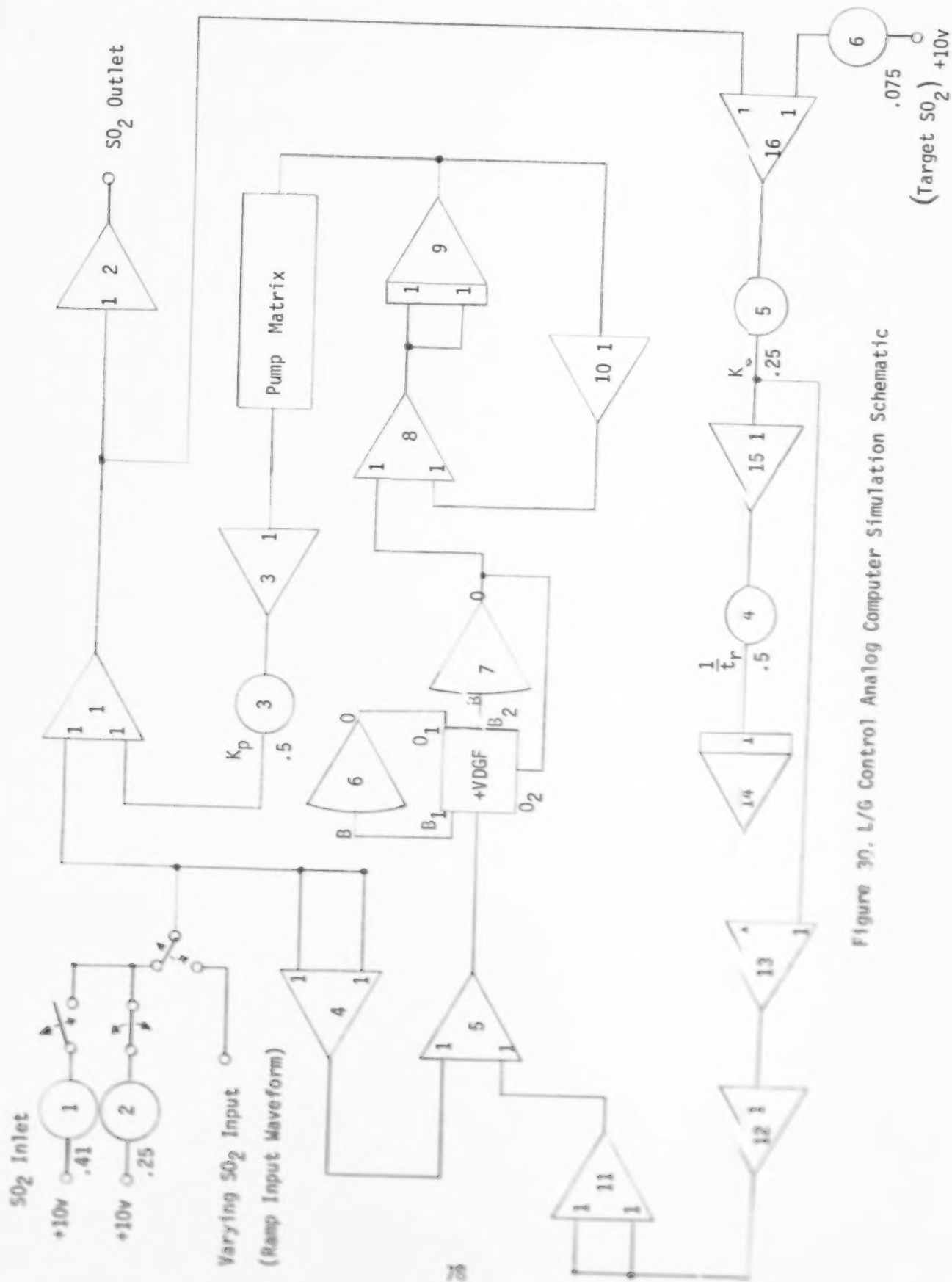


Figure 29. Absorber Feedforward L/G Control System

stable feedforward L/G control to reduce scrubber energy requirements. The addition of an organic acid slurry buffer, such as adipic acid, can provide additional energy savings by increasing SO_2 removal efficiency thereby permitting a reduced L/G ratio. This addition can easily be worked into these results, and predicts the furloughing of one 4000 liter per minute slurry pump for L/G requirements above 8000 LPM/MCMM with $\frac{1}{2}$ percent acid addition of the limestone feedrate.

Due to the faster time constants associated with the absorber reaction, the simulation time scaling for the L/G control studies corresponds to 1 second of computer run time per minute of the process. The load change information was also different and ranged from an SO_2 inlet of 450 ppm to 3000ppm. For the feedback controller which measures outlet SO_2 and adjusts the feedforward SO_2 exit target value, potentiometer 6 sets the target value, potentiometer 4 the controller integral term t_r , and potentiometer 5 the proportional band. The nonlinear feedforward law curve was approximated by the straight-line fit of $\text{L/G} = 0.46\text{SO}_2 - 14$ by amplifiers 6 and 7 and a variable diode function generator. Amplifiers 8, 10, and integrator 9 simulated the absorber time constant of $1/0.5S + 1$. The pump-matrix selector logic was implemented with four discretely biased differential comparators adjusted to implement the staircase equation in Figure 30 corresponding to

$$\sum_{i=1}^4 u(1 - \Delta \frac{\text{FFL} \cdot \text{SO}_2}{\text{L/G}}) \cdot \text{L/G}$$



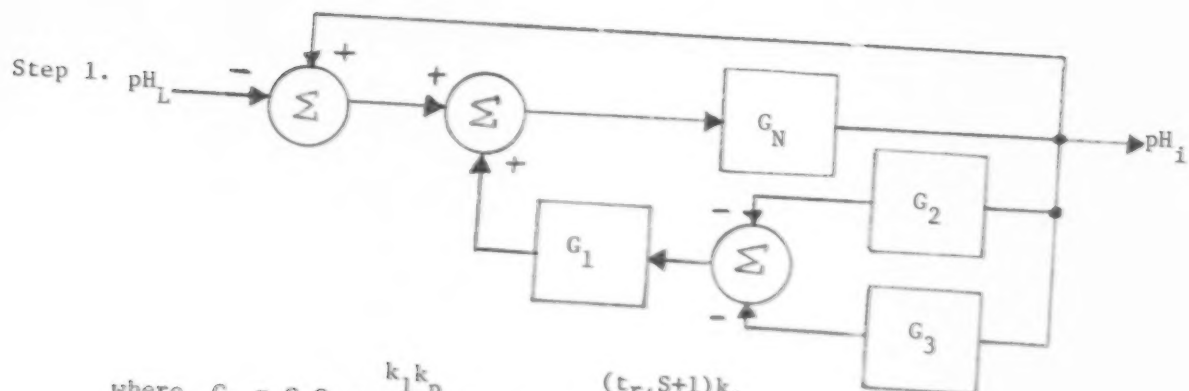
SECTION 12

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APPENDIX A



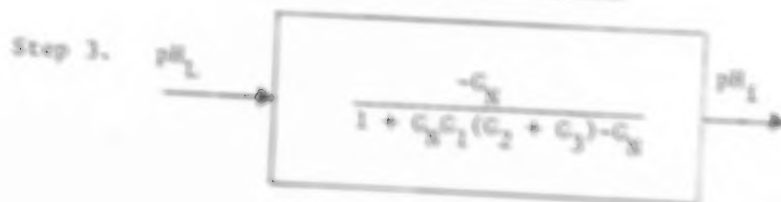
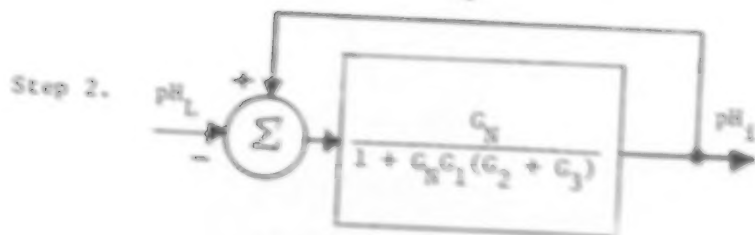
$$\text{where, } G_1 = G_d G_f = \frac{k_1 k_p}{1 - k_2 e^{-t_d s}} \cdot \frac{(t_{r4} s + 1) k_4}{(1 + k_4) t_{r4} s + 1}$$

$$G_N = \frac{1}{t_d s + 1}$$

$$G_2 = 0 \quad , \quad 5.4 < pH < 5.7$$

$$= \frac{k_o (t_{r_o} s + 1)}{t_{r_o} s} \quad , \quad \text{all other pH}$$

$$G_3 = \frac{pH_L k_Q}{t_{r3} s} = \frac{pH_L \cdot k_5}{t_{r3} s} \quad , \quad \text{where } k_5 = k_Q$$



Step 4.

$$\frac{pH_1(S)}{pH_L(S)} = \frac{-\frac{1}{\tau_d S+1}}{1 + \frac{1}{\tau_d S+1} \left[\frac{(\tau_{r4} S+1)k_4}{(1+k_4)\tau_{r4} S+1} \right] \left(\frac{k_p k_1}{1-k_2 e^{-\tau_d S}} \right) \left[\frac{(\tau_{r0} S+1)k_0}{\tau_{r0} S} + \frac{pH_L(S) \cdot k_5}{\tau_{r3} S} \right] - \frac{1}{\tau_d S+1}}$$

Step 5.

$$pH_1(S) = \frac{-pH_L(S)}{\left[\frac{(\tau_{r4} S+1)k_4}{(1+k_4)\tau_{r4} S+1} \right] \left(\frac{k_p k_1}{1-k_2 e^{-\tau_d S}} \right) \left[\frac{(\tau_{r0} S+1)k_0}{\tau_{r0} S} + \frac{pH_L(S) \cdot k_5}{\tau_{r3} S} \right] + \tau_d S}$$